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Cover Image: View of Bahrain Petroleum Co.'s (BAPCO's) Sitra refinery's expansion and modernization project. The nearly $7-B project—a major initiative of Bahrain's Vision 2030 program—will improve crude processing flexibility, increase energy efficiency, reduce emissions and enable BAPCO to produce ultra-low-sulfur transportation fuels.

HP Industry Perspectives

Innovative technologies for a sustainable future: IRPC returns as an in-person event

In late September, *Hydrocarbon Processing* will again host the International Refining and Petrochemical Conference (IRPC) as an in-person event—the technical symposium has been held virtually since early 2021. For more than 14 yr, the IRPC technical event has gathered key players in the global refining and petrochemicals sector to share knowledge, best practices and the latest technologies, services and equipment that are revolutionizing the processing industries.

The two-day event will be highlighted by timely keynote presentations, including topics on sustainability, net-zero goals for EPC companies and an outlook for the refining market from companies such as Dow Chemical, Shell, Technip Energies and Wood Mackenzie.

These keynote presentations will be followed by deep-dive technology tracks on the following topics: plant design, engineering and construction; process optimization, emerging technologies, process controls, instrumentation and automation; digital transformation; the circular economy; biofuels and alternative/renewable fuels; FCC/hydrocracking; alkylation; maintenance and reliability; and gas treatment/processing, among others.

IRPC will also feature bonus online content from Bharat Petroleum Corp. Ltd., Hindustan Petroleum Corp. Ltd., Kuwait Petroleum Corp., MyRechemical, Raffineria di Milazzo and Tupras, among others. These presentations will focus on topics such as advanced process control, carbon capture solutions, chemically recycling plastics, adhering to stringent diesel specifications and more.

IRPC is an ideal venue for hydrocarbon processing industry professionals to network with their peers. The latest conference survey of IRPC attendees included the following statistics: 86% of attendees held the job title of Manager-level or higher, 96% rated the conference as valuable and 96% stated that they would recommend IRPC to their colleagues.

IRPC 2022 will be held in Houston, Texas (U.S.) on September 26–27. Additional details about the event, access to the agenda, a view of the advisory board and pathways to register or sponsor can be found by visiting www.hpirpc.com. HP

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SUBSCRIPTIONS

Subscription price (includes both print and digital versions): One year $399, two years $679, three years $897. Airmail rate outside North America $175 additional a year. Single copies $35, prepaid.

Hydrocarbon Processing's Full Data Access subscription plan is priced at $1,995. This plan provides full access to all information and data *Hydrocarbon Processing* has to offer. It includes a print or digital version of the magazine, as well as full access to all posted articles (current and archived), process handbooks, the *HPI Market Data* book, Construction Boxscore Database project updates and more.

Because *Hydrocarbon Processing* is edited specifically to be of greatest value to people working in this specialized business, subscriptions are restricted to those engaged in the hydrocarbon processing industry, or service and supply company personnel connected thereto.

Hydrocarbon Processing is indexed by Applied Science & Technology Index, by Chemical Abstracts and by Engineering Index Inc. Microfilm copies available through University Microfilms, International, Ann Arbor, Mich. The full text of *Hydrocarbon Processing* is also available in electronic versions of the Business Periodicals Index.

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Hydrocarbon Processing (ISSN 0018-8190) is published monthly by Gulf Energy Information, 2 Greenway Plaza, Suite 1020, Houston, Texas 77046. Periodicals postage paid at Houston, Texas, and at additional mailing office. POSTMASTER: Send address changes to Hydrocarbon Processing, P.O. Box 2608, Houston, Texas 77252.

Publication Agreement Number 40034765 Printed in USA

Other Gulf Energy Information titles include: *Gas Processing™, H2Tech, Petroleum Economist©, World Oil®, Pipeline & Gas Journal* and *Underground Construction.*

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Although more than 170 yr old, the modern refining industry continues to evolve

As detailed in *Hydrocarbon Processing's* History of the HPI series, the modern oil refining industry traces its origin back nearly 170 yr. The industry has not only propelled much of the world to a higher standard of living, but has also created new products and technologies that have extended life expectancies, created modern societies and lifesaving equipment, and revolutionized the way people travel and communicate.

Due to oil demand, global refining capacity has increased to more than 102 MMbpd. Over the past several decades, these capital-intensive facilities have invested hundreds of billions of dollars in new units and technologies to increase clean fuels production, mitigate emissions, increase energy efficiency and incorporate new plants to provide products society demands.

According to *Hydrocarbon Processing's* Industry Market Outlook: 2022 Update webcast, the refining industry continues to hold the greatest market share vs. other processing industries; however, the sector has witnessed a slowdown in total active projects globally. Active refining projects have fallen since 2020, a direct effect of COVID-19 lockdown restrictions that led to fuel demand destruction globally. From 2020–2022, active refining projects have decreased year-over-year (y-o-y) by 27% and 7%, respectively. This decline contrasts with 2018–2020 when active refining projects increased y-o-y 11% and 10%, respectively (**FIG. 1**).

Moving forward. Many of today's refining projects are focused on producing low- and ultra-low-sulfur fuels while trying to mitigate carbon output. As the world shifts towards decarbonization, the refining industry has and continues to invest heavily in not only mitigating emissions from processing plants but also incorporating the latest technologies to produce sustainable fuels.

These major trends and the evolution in refining technologies have been the basis for technical coverage in the pages of *Hydrocarbon Processing* for the past 100 yr. They are also the primary reason why the publication is devoting an entire Special Focus to the novel techniques and processing technologies that can assist in not only decarbonizing oil processing but also lead to the green refinery of the future. HP

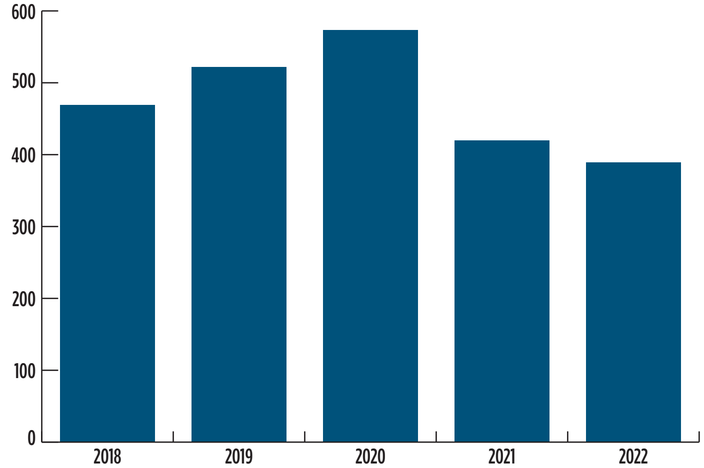

FIG. 1. Active refining projects globally, 2018–2022. Source: Global Energy Infrastructure database.

INSIDE THIS ISSUE

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IT/OT convergence: Efficiency and cybersecurity in the energy and natural resources sector

Information technology (IT) and operational technology (OT) environments have traditionally assumed distinct roles in industrial organizations. Separated both in purpose and in practice, IT and OT also differ in their culture and their norms around security and efficiency.

However, as new technologies are introduced to the OT environment, it is becoming necessary for the two environments to be converged. Without convergence, the new technologies and the data gathered will be vastly underutilized, making it difficult to justify the investments needed to modernize OT environments. Convergence requires organizations to bridge the gaps between IT and OT environments' people, processes and systems and to build a smarter, more secure network with high visibility to monitor and control both environments.

This article considers the challenges and opportunities and presents a path toward better convergence of people, systems and strategies between these two functions of an organization in the energy and natural resources sector.

Starting from behind. Transformational change is not just about execution or follow-up: it is also about preparation. IT/OT convergence requires the right preconditions—in both an organization's environment and culture—to be successful and lasting.

Though some of the world's most successful industrial, and energy and natural resources companies are well along the path of IT/OT convergence, many of their smaller or less ambitious peers are not. Each in their own way, systems, people and strategy are holding them back from starting or are limiting their progress with IT/OT convergence.

Systems. OT systems are usually quite a bit older than IT systems due to their capital intensive nature and safety protocols that prefer consistency to change. By design, newer OT environments are, for the most part, already prepared for convergence. These systems can support full automation, and remote monitoring and Internet of Things (IoT) devices can easily be integrated.

Taken alone, system age is not necessarily a hindrance to IT/OT convergence. An OT environment can be upgraded to support automation, monitoring and IoT devices. However, OT systems that use outdated equipment or are only patched on a minimal level are not secure. Many organizations do not secure these systems because of a desire to separate them from the internet and other networks. Securing OT systems is a prerequisite to IT/OT convergence.

Cybersecurity capabilities must be implemented to evaluate existing systems for threats and to continually monitor them in the future.

Another system prerequisite for IT/OT convergence is separation. This may seem counterintuitive, but if an IT and OT architecture are clearly separated prior to convergence, their eventual integration will be easier to accomplish. It is critical this is done at the beginning of the convergence project execution lifecycle rather than midway through it.

People. Preparing an organization's people and culture for IT/OT convergence is critical for success. In fact, it is often the first step. Building a strategy for convergence should start at the top.

Unfortunately, simply bringing the idea of IT/OT convergence to the board room is a struggle for some companies. When the long-term benefits of convergence are not fully understood, it is difficult for its proponents to articulate the need for convergence to management or the board.

Poor understanding also results in inaccurate evaluations of return on investment (ROI). To remedy these difficulties, it is important for organizations to include both IT and OT administrators in the evaluation of an IT/OT convergence proposal. Too often at industrial companies, IT administrators are left out of these conversations.

In the same vein, Chief Operating Officer's (COO's) opinions are often heard before the Chief Information Officer's (CIO's) opinions at industrial organizations. However, both executives have important contributions to make towards the planning of an IT/OT convergence strategy. Therefore, for organizations hoping to explore IT/OT convergence, CIOs should be empowered on management teams and in board rooms and not seen as a second fiddle to COOs.

Cybersecurity should be integral to an IT/OT convergence project from its inception. Responsibility for IT/OT convergence cybersecurity will often fall on a Chief Information Security Officer (CISO). In many cases, CISOs are brought into a convergence project in the middle or even late stages, which can mean systems and processes are not integrated securely. When CISOs are included in the planning stages, they can identify the areas most vulnerable to threats and develop the plans to secure them early on.

Strategy. A good practice for organizations considering IT/OT convergence is to clearly define what use cases they want to solve by converging environments. Do you want your systems to operate more efficiently, deliver faster insight to leadership on production data and analysis, or do you want to improve maintenance and servicing of equipment?

If organizations fail to define clear objectives for IT/OT convergence, they may also fail to implement the right processes and tools for their specific environment, or they may end up with diffuse goals that cannot be accomplished. Both outcomes diminish the effectiveness of convergence and will reduce long-term morale and buying around convergence.

However, in defining a convergence strategy and deliverables, organizations should allow for flexibility. Especially within the cybersecurity realm, threats and protections are rapidly evolving—this should be addressed by infusing flexibility into your strategy. While it is easier said than done, finding a middle ground allows for adaptation when inevitable difficulties are encountered while also maintaining clear project goals and timelines.

Process and workflow convergence.

Process and workflow convergence are integral to a broader IT/OT convergence plan. Introducing new technology to a converged system without adapting processes and workflows to the new system will not deliver the desired business benefit. Therefore, before organizations can begin converging processes, it is important to recognize the reasons why IT and OT processes are different. Once the "why" is identified, an organization can look at how processes can be converged without compromising the effectiveness or safety of an individual IT or OT process. IT and OT processes and workflows have always been different—for good reason.

Underlying nearly every OT process is safety. Consistency is safety's best friend, and as such, OT systems favor legacy systems and processes—change is rare in applications and infrastructure and there must be a good reason to do so. IT processes are quite the opposite. To achieve greater process efficiency or to adapt to evolving risks, IT processes encourage updates and adaptation.

Both philosophies support cybersecurity in their own way. IT processes can be secure because they are constantly upgraded to address vulnerabilities as they emerge from the wild and use new cyber defenses. OT processes can be secure due to their strict adherence to procedures and regimented control of change.

Before entering a converged environment, both IT and OT cybersecurity processes and workflows must be adapted. A great deal can be learned from one another.

Stricter attitudes towards processes and a culture of safety—hallmarks of the OT environment—should be adopted within IT. Now that their work is more directly integrated with manufacturing or production systems and the personnel physically operating them, IT administrators must recognize the elevated stakes associated with cybersecurity. The resulting cultural change within IT should better prepare IT processes and workflows for convergence.

Conversely, OT processes and workflows should be adapted to fit a more regular schedule of updates. This approach is necessary to support cybersecurity in a converged environment that contains more connected devices and potential vulnerabilities. IT administrators are acquainted with this approach and their expertise should be utilized when designing new OT processes, systems and capabilities to support convergence.

OT processes and workflows can be prepared for convergence—and supported during and after convergence—by an OT security center for excellence. A center for excellence is important for training OT administrators and engineers to protect critical infrastructure as new vulnerabilities are introduced in IT/OT convergence. Administrators can learn through hands-on cybersecurity exercises that replicate real-world risks. In turn, these engineers can tailor their workflows around a new understanding of cybersecurity in a converged environment.

Management- and board-level thinking. People—rather than systems or technologies—may be the most important factor for IT/OT convergence success. However, while much attention is paid to the actual practitioners of IT and OT systems (i.e., those with their hands on the physical levers), often lost in the conversation are the people at the top of an organization: those at the management and board level.

In addition to determining an organization's budget or goals, the management team and board determine culture. Two key aspects to a culture that supports IT/OT convergence are understanding and open-mindedness. Though these may sound better suited for marriage counseling than for an industrial operation, they are practical for both.

Increased understanding of each other's systems and processes is critical for preparing IT and OT teams for convergence. Such understanding will improve both the efficiency and safety of newly converged processes like data transfer and system monitoring.

Open-mindedness—to one's IT or OT counterparts—will ease the acceptance of new shared protocols or processes resulting from IT/OT convergence. If administrators are aware of the unique needs of an IT or OT system, they are more likely to buy in to the plan and accept compromises to how they think the operation should be run. There are additional practical steps organizations can make to their management team and board to support IT/OT convergence.

The CIO does not sit on the board for many industrial organizations. This fact is a relic of an old way of thinking—that operations trump all, especially IT. This broadly needs to change, but particularly for organizations preparing for IT/OT convergence. IT must have a seat on the board so that their needs and insights are addressed for convergence plans.

This approach can be applied more broadly to management with a leveling of opinions and responsibilities. The COO's opinion at an industrial organization should not necessarily be more important than that of the CIO. Chief Financial Officers (CFOs) should fully understand the risks and benefits of IT/OT convergence so that their decision-making supports the effort. Full management buy-in is only possible when everyone's opinion is valued.

Distributing responsibilities between IT and OT. Somewhat counterintuitively, a key for IT/OT convergence success is knowing which processes to keep separate. Within this, distributing cybersecurity responsibilities between IT and OT teams is critical to ensuring convergence does not make an organization more vulnerable than before.

CISOs have become increasingly important for industrial companies. However, within large, diverse organizations, CISOs' differing responsibilities for different divisions can become burdensome and even dangerous if their holistic remit prevents the separation of cybersecurity responsibilities. A way to avoid unnecessary vulnerabilities resulting from an overstretched CISO is to have two CISO-like roles—one each for IT and OT—in addi-

tion to an executive CISO to which they both report. The benefit of having an IT CISO and an OT CISO is to support a cybersecurity separation strategy meant to prevent a cyberattack moving from the IT to the OT environment, or vice versa.

A two-CISO strategy is illustrative of a broader strategy to distribute cybersecurity responsibilities in a manner that prevents attacks moving from one environment to another. The administrators responsible for the IT firewall should not also oversee OT's firewall. This makes sense on a practical level—two teams of administrators better separate systems and controls—and on a philosophical level, using one line of thinking to create and maintain two firewalls leaves a system more hackable than using two lines of thinking.

A prerequisite for effective distribution of cybersecurity responsibilities is complete asset visibility. The ability to monitor assets means first getting and maintaining visibility of critical assets, then using risk-based methods when planning and implementing protective security monitoring.

Enhancing capabilities. Broadly-tested reference architectures and more-specialized solutions relating to data management and storage can enhance an organization's IT/OT convergence efforts.

Network security reference architectures and zero trust. Establishing a reference architecture for network security during IT/OT convergence will yield a multitude of benefits. Reference architectures help all stakeholders collaborate and communicate effectively throughout a process—a notoriously tricky task between IT and OT teams. With a reference architecture in place to anticipate network security questions that may arise and provide objective guidance to answer them, the subjective differences in opinion between IT and OT will be minimized.

ISA/IEC 62443 standards can be a valuable resource for organizations undergoing IT/OT convergence. The standards address security issues unique to connected OT systems. Among the security-related convergence tasks that IEC 62443 addresses are:

- OT cyber risk assessments
- Building cybersecurity management teams
- Patching and other protective controls
- Segmenting and securing network zones and conduits
- Creating processes and governance
- Establishing appropriate roles and responsibilities for users or resources.

Additionally, choosing and managing the right connected devices for a unique OT system are important for securing networks for IT/OT convergence. IEC 62443 addresses this and provides guidance to improve existing processes for technology project scoping, vendor selection and procurement. It also contains a set of prescriptive requirements and processes for secure product development lifecycles fit for a connected OT environment.

Traditional approaches for securing OT networks have long been dependent on maintaining the separation of industrial applications from IT networks. However, for organizations undergoing IT/OT convergence, traditional separa-

tion approaches are insufficient.

Zero trust takes traditional separation approaches to a much more granular level. Rather than building a wall around the OT network, it assumes that every user, device, system or connection is already compromised (by default) whether they are inside or outside of the network. The goal of zero trust is to reduce the attack surface and prevent lateral movement with segmentation or micro-segmentation. If a breach occurs, then an intruder cannot easily access other systems or sensitive data by moving laterally.

Data management. One of the greatest benefits of IT/OT convergence is the generation of previously untapped data. These data can be used to enhance the efficiency of an operation, increase security or conduct predictive maintenance. However, without effective data management practices, much of these data's value is unrealized.

While many organizations manage their data by sending it to the cloud for storage or processing, this method does not allow organizations to tap into the full potential of industrial IoT capabilities. Processing a large amount of data directly in the could requires lengthy response times and a large amount of bandwidth. An alternative solution is edge computing.

Edge computing brings data storage and processing closer to the sources of data. Edge computing processes happen on the floor of an industrial facility, meaning only some data must be sent to the cloud for long-term storage or distribution to a different network. Bringing analysis closer to home improves response time and bandwidth consumption. It also decreases data distribution and the inherent risk of that data being stolen or damaged during data transfer.

Data management should also encompass security. In addition to data storage and processing methods like edge computing, practicing selective sharing ensures that data is accessed only as it should be. Specific data might flow one way, back and forth, or not at all. Limiting the flow of data makes it less vulnerable to threats.

Evolving cybersecurity risk factors.

OT systems—rather than IT systems—are usually carrying an organization's most critical business processes. Therefore, OT system downtime can be more costly for an organization than IT system downtime. Logic suggests that OT system security should be more important for a business than IT system security.

However, the paradox is that most companies invest a lot more money into securing their IT systems than into securing their OT systems. This investment paradigm does not necessarily need to be flipped, but a more balanced approach to cybersecurity planning and investment must be taken during IT/OT convergence.

IT/OT convergence improves cybersecurity. IT/OT convergence allows for broader monitoring and threat detection. Prior to convergence, organizations will likely have a limited view into the architecture and services running on their OT system. Convergence with IT allows for passive monitoring and vast data gathering. In analyzing these data, profiling techniques can be used to learn more about the tendencies of a system. Once data patterns are better understood, abnormal behaviors can be quickly detected with no interruption to the industrial process.

Broader data gathering has benefits beyond threat detection. New datasets can be fused with data sets from other systems across the OT system (e.g., closed-circuit television or building management systems) to allow for more holistic monitoring and better utilization of data.

Factory maintenance can also be improved through IT/OT convergence. Most organizations rely on preventive maintenance to decrease the likelihood of system downtime and costly repairs. IT/OT convergence allows organizations to shift from preventive maintenance to predictive maintenance. Predictive maintenance technologies gather real-time data from OT systems and can spot situations that might lead to equipment failure. In doing so, targeted repairs can be made prior to failures—this method is often less-costly than preventive maintenance, which takes a broader approach to repairs.

Organizations must bear in mind that the cybersecurity advantages of IT/OT convergence are limited by the type and age of an organization's OT system. Older supervisory control and data acquisition (SCADA) systems often do not allow for the broader monitoring, threat detection or data gathering that is necessary to build a more secure ecosystem. OT systems that employ more IoT and connected devices have the access points for monitoring and data gathering and are more suited for the cybersecurity benefits of IT/OT convergence.

IT/OT convergence can open cybersecurity vulnerabilities. While IT/OT convergence can allow for more robust cybersecurity, it also exposes industrial control systems (ICS), process control systems and other operational technology to malware attacks, hacktivism, employee sabotage and other security risks that previously affected only corporate IT systems.

IT/OT convergence involves integrating IP-connected devices to an OT system. These additional access points make an OT system more vulnerable to internet-based threats. Many of these connected devices use consumer-grade technology and are from multiple vendors—both qualities that make them susceptible to cyberattacks.

Often, the trickiest management aspect of IT/OT convergence is organizational change. Systems do not have opinions, people do. The knowledge and cultural differences between IT and OT employees can result in vulnerabilities that can be exploited by cyber attackers. One of the key differences between the IT mindset and OT mindset is the reluctance to or acceptance of system upgrades. While IT administrators constantly update their systems for optimization and security, OT admins are more likely to stick with legacy systems because of their reliability.

A mindset clash between IT and OT administrators can lead to poor system integration. This results not only in sub-par IT/OT convergence outcomes, but also in system vulnerabilities.

Vulnerabilities created by hosting different OEMs in a single environment. Managing multiple original equipment manufacturers (OEMs) in a single environment is difficult from an operational perspective since most OEMs' systems were not designed to work optimally with systems from different vendors. Such system idiosyncrasies can also impact the safety of an environment. As a result, system administrators must identify vulnerabilities and patch where necessary.

In a system containing devices and subsystems from multiple OEMs, it is crucial for an organization to define and limit the paths for data transfer. Data transfer is crucial for successful IT/OT convergence; without it, systems will be poorly integrated, and convergence will not yield the desired system optimization. However, data

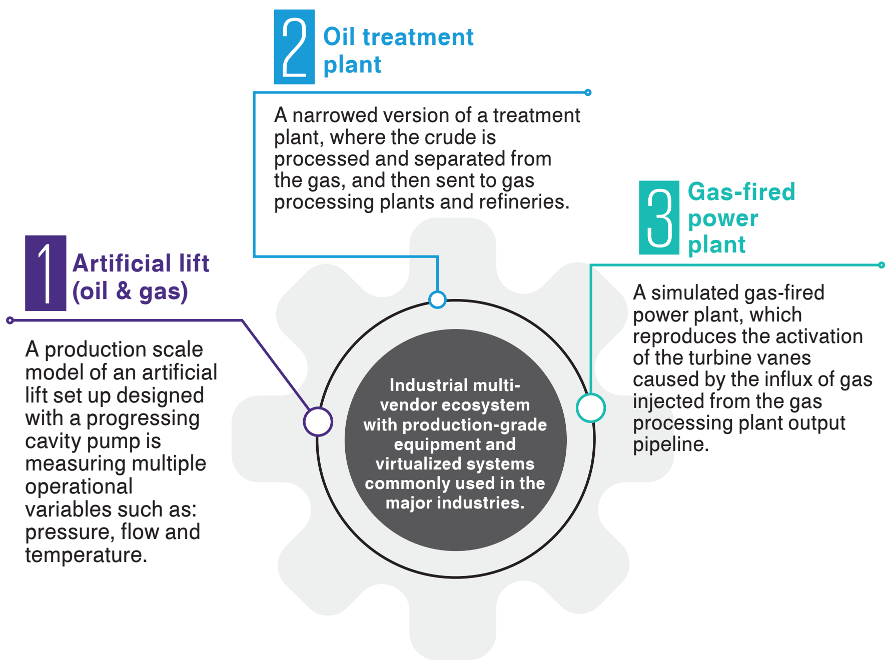

FIG. 1. OT/ICS cyber range labs examples.

transfer should be limited to paths where it can be useful. If two devices—especially from different OEMs—have no need to transfer data, they should not.

However, there are differences in opinion about the safety of hosting multiple OEMs in a single environment. Some administrators see multiple OEMs as a protection against attackers moving easily from one system to another. In their opinion, having a single OEM leaves an organization vulnerable to having their entire environment compromised.

Mitigating zero-day attack risks during and after IT/OT convergence. By virtue of their definition, zero-day attacks are impossible to predict. However, they are possible to defend against. In converging IT and OT systems, an organization is turning a closed environment into an open environment. This creates new vulnerabilities to zero-day attacks that must be addressed.

Micro-segmentation is one practice employed by organizations to mitigate the impact of a successful zero-day attack. In a converged IT/OT environment, traditional security controls such as virtual local area network (VLAN) systems are insufficient protection from a zero-day attack. Micro-segmentation provides a more targeted control over network traffic by further partitioning the VLAN and adding security for each partition.[1]

Furthermore, micro-segmentation can be tailored to a specific operating environment and for specific types of network traffic. This allows for the limitation of certain traffic without significantly minimizing the optimizing benefits of IT/OT convergence.

On a fundamental level, organizations can mitigate zero-day attack risk by implementing resilient-by-design principles before IT/OT convergence. Instituting resiliency in your organization's people, processes and systems prior to convergence will also minimize the impact of a zero-day attack, should it occur.

OT/ICS cyber range labs. Simulating a cyberattack in an OT environment is a difficult process. While many IT cybersecurity professionals are well-versed in methods like penetration testing and red teaming to test the strength of their cyber defenses, OT cybersecurity professionals struggle to find widely-used practices to simulate OT cyberattacks.

To bring OT simulation efforts up to par with IT, the authors' company has created OT/ICS cyber range labs using production grade equipment to simulate scale model versions of industrial processes. The labs can be used to establish secure remote connections through the authors' company's infrastructure to perform the following:

- Hands-on training sessions
- Cyberattack simulations
- Proof-of-concepts (PoCs)
- Industrial cybersecurity-related research.

Lab simulation examples are shown in **FIG. 1.**

No two industrial environments are the same. Resultingly, training is often hyper-specialized for an organization's specific set of OT environments and systems. In the virtual environment, OT/ICS labs can be tailored to match the specific training needs of an industrial operation.

These virtual labs can be built to replicate an organization's IT and OT environments by connecting proprietary devices and virtualizing OT components. This enables IT and OT professionals to cross-train their incident response strategies until mastery. The following are examples of lab use cases:

1. **OT cybersecurity hands-on training sessions:** Labs can be used to perform online training sessions with production-grade equipment to learn OT/ICS cybersecurity concepts. Trainees will become acquainted with performing their daily tasks in a cybersafe way.
2. **OT cybersecurity tool demos:** Hands-on demos of cybersecurity tools help OT employees better understand the technical capabilities of the tools and evaluate their use in multiple scenarios.
3. **OT cyberattack demos:** In the labs, multiple cyberattack scenarios can be applied to the industrial environment to show how OT assets function when crippled. The scenarios help reveal vulnerabilities and what equipment might be compromised in an attack.
4. **OT cybersecurity remediation testing:** Prior to implementation of a cybersecurity remediation plan, labs can be used to perform testing procedures to review the effectiveness and reliance of a workaround in a similar architecture. HP

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JASON URSO, Chief Technology Officer,
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The pathway towards autonomy: How process automation is building the future plant

JASON URSO is Chief Technology Officer, Honeywell Process Solutions, a leading provider of industrial automation and digitization technology, smart devices, thermal process solutions and advanced software that help operators around the world improve their efficiency, sustainability, safety and throughput.

Hydrocarbon Processing (*HP*) sat down with Jason Urso (JU), Chief Technology Officer, Honeywell Process Solutions, to discuss the current state of the process automation industry, how new technologies are enhancing efficiency and safety, and what the future holds for automation.

HP: What does the future plant look like and how can process industries get there?

JU: Major industrial producers today are focused on sustainability, productivity, safety and compliance. This involves environmental-focused initiatives like helping to reduce carbon footprints and improving safety of the plant, personnel and equipment. It also comprises new areas, such as ensuring the right level of talent are available due to the challenges associated with an aging workforce and reluctance from the next generation of workers to seek employment in oil and gas-related industries. Industrial autonomous operations provide a future where industrial facilities can continue to improve performance in these areas. By moving to a higher degree of autonomy, industrial operators can improve safety and environmental performance, increase efficiency over historical levels, improve reliability and attract highly skilled talent.

The future physical plant looks similar from a distance—the same pipes, columns, etc., are being used to process product. However, a significantly higher number of sensors are being employed with technology (e.g., drones and robots) to bring in data more frequently than can be accomplished by humans. New talent will be trained with technologies like augmented reality (AR) coupled with digital twins of processes and plants where they are able to learn how to physically respond to events that were previously taught on paper due to their hazardous nature or the infrequent occurrence of the operation. Systems in the plant must become increasingly self-diagnostic and eventually self-healing. As industry relies more on sensors, personnel need to consistently know their status in case a sensor malfunctions. The future of highly autonomous plants brings a higher degree of awareness in both its processes and systems.

HP: How have current automation technology and software solutions impacted process industries today?

JU: Automation technologies like digital twins, advanced process control, historian and control systems are resilient and easy to engineer, maintain and upgrade—they form the basic backbone of industrial operations. Advancements in these technologies enable a cybersecure data exchange to the cloud and apply smart analytics, artificial intelligence (AI) and machine learning (ML) to the data to gain actionable insights. This is the path forward for the process industries to become more efficient, safer and more compliant. Some of the key drivers and enablers for autonomous technology in industries today are the relatively low cost and availability of sensors coupled with the advancement of data science and analytics to turn data into information. By adding sensors into the process, we can apply digital eyes (video), ears (sound), feel (e.g., vibration and heat) and smell (e.g., emissions). While these sensors are not purposed to replace human monitoring, they offer an advantage; they can be applied continuously or at least more frequently than what a person can do. Getting the data from these devices enables personnel to leverage data analytics to analyze what has been detected to a higher degree than ever before. When people sense something through routines like operator rounds, a report or log of the observation is created, but it is a qualitative account. Using sensors enables workers to log quantitative values and evaluate trends vs. simply comments or observations. By analyzing and correlating trends of quantitative data, plant personnel can detect and prevent undesirable events. For example, by detecting increasing levels of vibration, heat or sound, workers can detect mechanical issues long before equipment reaches a point of failure.

HP: How will autonomous operations improve plant safety, reliability and efficiency?

JU: Autonomous operations can look at multiple aspects of plant operation

FIG. 1. Industrial refining setting.

and optimize it based on the entire value chain, from incoming raw material, energy used (green energy), emissions compliance and productivity to final finished product output.

As industry detects abnormalities in processes faster, personnel can take many more planned actions prior to a failure. If we run plant equipment to failure, we risk having challenges during startup, shutdown and, ultimately, we can jeopardize the stability of the operation. When abnormalities, or even variables trending away from normal operations, are detected early, planned actions can be taken. Personnel can switch to spare equipment, have required parts and expertise on-hand for anticipated maintenance, make repairs faster and have all equipment back to a normal state much faster when challenges are anticipated.

Beyond managing these abnormal situations, autonomous operations enable a higher degree of optimization. Leveraging improved data analytics, AI and ML can optimize industry operations. We can minimally duplicate what the best human operators do and make every operator the best operator at the facility. Our organization has been a member of the Abnormal Situation Management consortium for many years. Here, member industrial companies lead studies and work with other manufacturers to reduce human error and help to make operators more effective. Since our company is both a technology provider and a manufacturer of chemicals in our own plants, we have a unique capability and desire to improve operator effectiveness for our customers' and our own plants.

HP: How has the industrial evolution of control technology benefited plants and facilities today?

JU: Processing plants today operate significantly safer, more efficiently and more reliably than ever before. This has been possible largely through automation. The industry has progressed from old technologies when plants were controlled by pneumatic controllers. When digital controls became available, industrial producers saw significant improvements. Processes were controlled more precisely than ever before. Data was captured and stored electronically vs. by paper chart recorders. As processes were controlled better, industry was able to optimize—the next level of advancement came with advanced process control (APC) where control technology not only controlled input variables like temperature, pressure or flow, but also controlled output variables. With this advancement, the process industry was able to control quality parameters and optimize production across facilities. This fueled the development of soft sensors and inferential matters that were early applications of AI. In turn, this enabled prediction and control of outputs like quality parameters based on mathematical calculations and correlations without the need for many instruments and analyzers. This evolution has established a strong foundation that continues to develop into industrial autonomous solutions today.

HP: Have you noticed an increase in autonomous solutions implemented by industrial leaders?

JU: We absolutely see an increase in autonomous solutions implemented by industrial leaders. Most industries are challenged to improve the safety, reliability and efficiency of their processes, and because autonomous solutions can provide benefits in this area, we see interest from most of our customers. As with any technology advancement, we also see leaders in adoption while others take a wait and see approach. As the leading industrial operators are experiencing significant benefits in their projects, the wait and see segment of customers will start to catch up. Autonomous solutions increase the ability to train new talent faster as well as attract higher skilled talent with an appetite for technology; those early adopting customers are staffing with the best and brightest.

As more autonomous technology is applied to processes, we are better able to connect remote expertise to support producing sites—this has enabled the use of remote hubs of talent for monitoring, support and surveillance of enterprises. This area developed rapidly in the last two years as part of our customers' pandemic response. By connecting support remotely, industrial employers were able to minimize exposure of their site-based workforce, as well as mitigate long quarantine periods for those that would have historically travelled to sites. As this culture of remote support grew, the automation industry quickly expanded the use of remote technology. We are now able to migrate customer automation systems from older versions of software to current versions using remote connectivity. We can also provide support to commission new startups without having engineers travel to the construction site, and we can test and prove new installations with remote engineers that have live connections and video feeds to the industrial site.

HP: Are there specific controls and remote solutions that have the flexibility to fit any plant or facility manager's custom needs?

JU: There are a few areas of autonomous technology that not only fit any plant's needs but that are becoming necessary to most of them. First, anything that is done must be done with the highest levels of cybersecurity—it is important that both the IT (information technology) and OT (operational technology) digital infrastructure are architected and protected with adequate cybersecurity protection. Once a cyber infrastructure is created, the facility can enable remote connections. By allowing remote views or even control of processes, most plants can improve their offsite support. From simply enabling on-call weekend or after-hours support, empowering leadership to see what is being seen inside the plant or connecting to corporate-level technical teams, remote connectivity provides significant benefits. Once remote connectivity is established and initial benefits experienced, some customers progress to remote operations. By creating control centers outside of hazardous process areas or in remote geographies, plants can establish higher levels of safety and improve access to talent.

HP: What steps should a plant/facility take to achieve industrial autonomy? Is there

a roadmap a plant manager/ supervisor can follow?

JU: Our company has developed a 5-stage maturity level map to identify where a plant or facility are on their journey to autonomous operations. These include the following:

- **Level 1:** Controlled and optimized
- **Level 2:** Intelligent plant
- **Level 3:** Remote and integrated operations
- **Level 4:** Resilient operations
- **Level 5:** Autonomous operations.

A facility must first identify where in this maturity level map they are and then they can work with a technology provider to develop the roadmap to autonomy. The journey to autonomous operation is not a one-size-fits-all approach. The roadmap should start with a strong plan for change management. We have seen examples where technology application without change management has created an end state of lower efficiency than the starting point. The desired results must be mapped with technology and change management processes that enable the result. Our organization has staffed a team of process change consultants, as well as partnered with other companies to facilitate our customers' autonomous journey. We have created standard deployment guidelines for industry verticals that describe available technologies and typical benefits experienced by each industry. By applying a consultative approach with our customers, we can participate in a discovery process and share which applications have been proven high-value in their industry.

***HP*: How does an autonomous operation help plants reach their sustainability goals and reduce their carbon footprint?**

JU: Autonomous operation can enable sustainability in several ways; we have already touched on how it enables a sustainable workforce. Operating autonomously can also have significant impact on carbon footprint. For example, by implementing autonomous technologies that measure emissions from the process, industrial operators can quickly detect process emissions and incidents like smoking flares and methane emissions. Our organization's new emissions control and reduction initiative allows customers in a wide range of areas to detect the precise location of methane leaks through a combination of detection and gas cloud imaging cameras, translating the information into comprehensive data analytics and trends. Customers will then be able to use this data to quickly identify leaks, increasing productivity and maintaining legislative compliance. This kind of technology can also help prevent the root cause of emissions by reducing process upsets and eliminating unplanned shutdown/startup activity. In addition to beneficial technologies such as these, our company has the capability to provide other solutions to reduce carbon footprint: technology to produce renewable fuels, game-changing battery technology for renewable energy storage and environmentally friendly refrigerants for processes, improving numerous areas that positively impact our customers' carbon footprint. HP

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Column revamp using packings and internal modeling and revamping technology

The challenge of frequent plugging in vent gas scrubbers is resolved with the highest degree of maleic anhydride (MAN) product recovery. This article describes the details of a problem in a packed bed column, the troubleshooting approach, an overall revamp study, features of the packing products supplied by the author's company and the benefits achieved by the authors' company.

The vent gas scrubber is a packed bed column equipped with a conventional structure packing bed. The column operates at a moderately elevated temperature and a pressure of 25 mbara. The lean oil solvent is fed to the top of the vent scrubber to absorb the MAN and the non-condensable vapor from the vent scrubber passes to the vent header. A rich oil stream from the bottom of the vent scrubber is sent to another column to recover the MAN.

A generalized schematic simulation scheme showing the packed bed arrangement is illustrated in **FIG. 1.**

The vent gas scrubber was experiencing frequent plugging of the column internals. At the start of operation, the column pressure drop was 5 mbar, but increased to as high as 20 mbar within 3 mos–4 mos. Due to the process nature of the application, deposits were detected on the packing during plant operation. This resulted in the plugging of the packed bed and eventually required column packing washing once every 3 mos–4 mos.

A simulation was performed for the column, and vapor liquid traffic hydraulics results were shared with the author's company, which validated the hydraulics calculation and proposed a next-generation proprietary structured packing[a] and a high-performance proprietary distributor[b] to resolve the problem.

Historic plant data.

The authors' companies studied the actual column pressure drop and plant data to analyze and understand the problem. From plant operation data, it was observed that the increase in pressure drop across the column was gradual; within 3 mos of operation, the column required solvent washing to remove the plugging material from the bed. The packing and internals were inspected during a scheduled shutdown to determine the reason for this high pressure drop. The internals were found to be intact, but severe fouling was seen by strong disposition of acidic materials due to the nature of the process.

The hydrolysis of MAN by water forms maleic acid as part of the process, and maleic acid is thermally converted to fumaric acid. The formation of fumaric acid accounts for most difficulties in the column high pressure drop due to blockage of the equipment and surface fouling.

The use of computational fluid dynamics (CFD) revealed that the shape of the existing packing type had supported the accumulation of liquid/solids at the junction of two subsequent structure packing layers due to the abrupt change of direction, which is detrimental for lower pressure drop.

FIG. 2 shows the existing packing pattern causing material accumulation at the packing junction. The effect of pressure

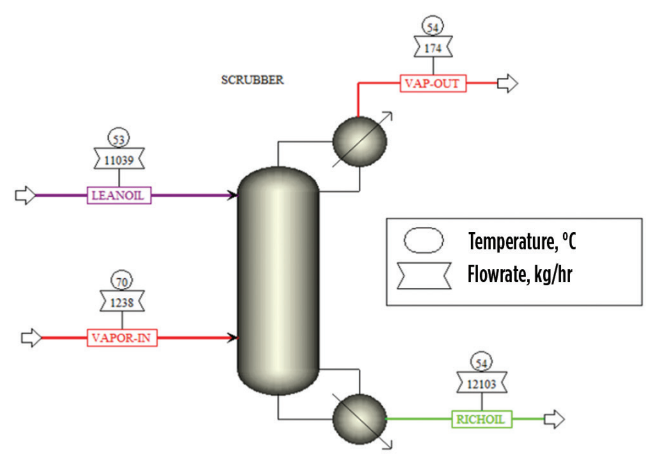

FIG. 1. Process flow diagram. Source: AspenTech.

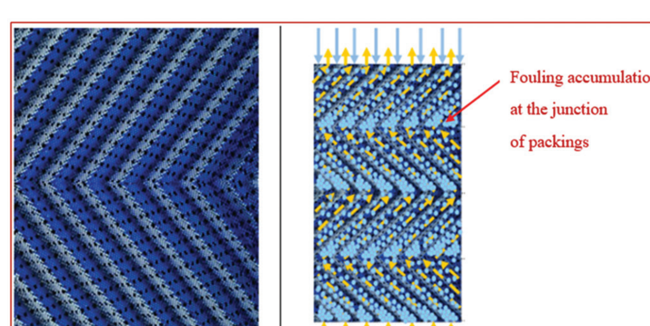

FIG. 2. CFD revealed that the existing packing shape supported the accumulation of liquid/solids at the junction of two subsequent structure packing layers.

drop on MAN recovery is directly proportional. The column operational trendline clearly specified that a low column pressure drop provides improved MAN recovery.

Most column revamps necessitate the evaluation and modification of ancillary equipment. However, associated equipment such as condensers, receivers, pumps, valves and instrumentation may remain as they are.

TROUBLESHOOTING APPROACH

In any unit operation, future modifications without a basic engineering study are a risky adventure. Simulation provided by an industry leader was used due to the company's widespread use of 40-yr databases of tuned models with specific vapor-liquid equilibrium (VLE) packages. These databases are important when designing distillation columns and are one of the key reasons that many distillation experts use the process simulation software[c]. In this case, a rigorous distillation calculation was performed on the existing column using the actual temperature, pressures, compositions, etc., of the column to accurately match the original heat and material balance and tune the steady-state model.

TABLE 1. Hydraulic performance comparison of old packing vs. proprietary structured packing[a]

Key factors	Existing packing	Proprietary structured packing[a]
Pressure drop at plant match load	2.6 mbar	< 1.5 mbar
% capacity	79.4	62.7
At 120% of plant match load	Packing is flooding	< 6 mbar

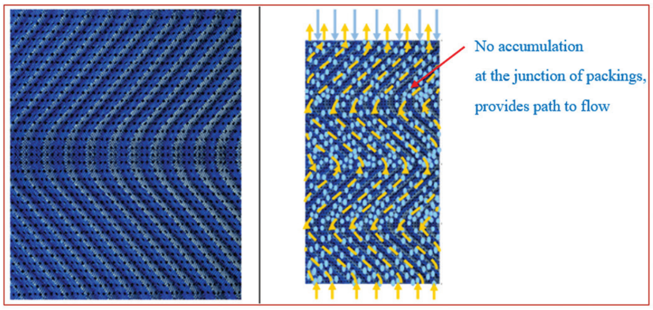

FIG. 3. The flow pattern of the proposed proprietary structured packing[a].

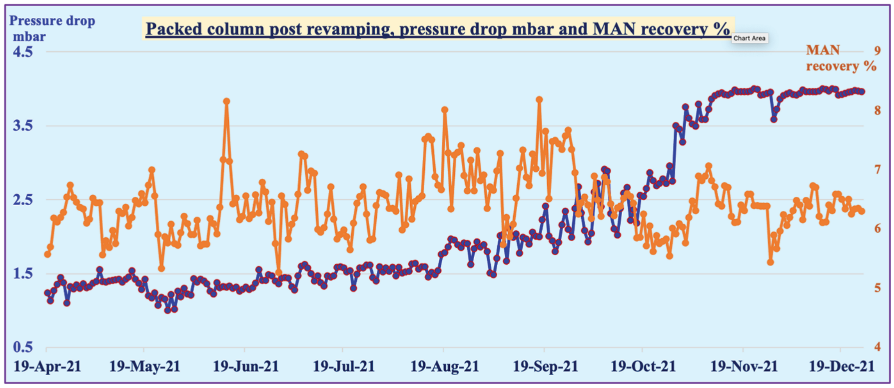

FIG. 4. Pressure drop and MAN recovery post revamp.

Rating the existing column. With the two tuned models, the column was rated for the new conditions. The rating determined that the existing equipment could be used in the new hydraulics operation. The hydraulic calculations were then evaluated, and the column capacity was reviewed using an analysis of simulated internal vapor and liquid flows. With the original heat and material balances, the VLE data and packed column efficiencies were checked. The existing vent gas column was rated for the correct number of transit units (NTS), reflux-to-stages ratio and packing hydraulics.

A plant match case simulation was performed with the objective of determining whether lower surface area packing should be used to improve column run length. However, the simulation showed significant loss of MAN from the top due to reduced NTS. Therefore, it was decided to not use the lower surface area packing.

Furthermore, from the author's company's hydraulics calculation of equivalent existing packing, it was observed that the existing packing bottom section was operating near to flood (79%) in a clean condition. This means that as soon as fouling starts in the packing, the surface area and overall cross-sectional area of the packing are reduced to the extent of the fouling, resulting in the packed bed operating under flood condition. As soon as the packing flooded, the pressure drop across the packing was increased exponentially.

Application of structured packing[a]. To resolve the structure packing capacity limitation and to reduce the overall pressure drop across the bed, the author's company checked the hydraulics of its latest generation structured packing[a]. The pressure drop across the packing is much lower and the maximum capacity can be extended up to 20%–30%. The new style of packing incorporates a patented modification to the lower and upper end of each packing element. The new packing's internals were verified with various operational variables/changes to meet the desired targets of the highest degree of MAN recovery by 6.5 wt%. During simulation, various options (pressure changes, temperature reflux rate changes, optimizing the feed, etc.) were studied.

For the revamp, different options were also studied, including random packings. The pros and cons of each option were identified and compared; in this service, however, random packing seems inapplicable due to the increased settlement of fouling materials and eventual high pres-

sure drop caused by liquid hold-up in the random packings.

Most column revamps necessitate the evaluation and modification of ancillary equipment. However, associated equipment such as condensers, receivers, pumps, valves and instrumentation may remain as they are. With the latest generation structured packing[a] with enhanced geometric structure, the pressure drop across the packing is much lower and the maximum capacity can be extended up to 20% compared to conventional structure packing in a simulation study. The new packing has a smooth curved shape at the junction of two packing layers, lowering the accumulation of liquid/solid at this point and resulting in high packing capacity. The corrugation angle in relation to the vertical is gradually reduced to zero at both ends of each sheet, as seen in **FIG. 3.**

This design modification of the corrugation angle causes smooth and steady changes of flow direction, as opposed to abrupt direction changes found in conventional structure packing. The result is a reduction in pressure drop and reduced shear force between the gas and liquid phase, and a reduction of gas velocity.

Hydraulics calculations of a packed bed based on plant match conditions showed that the proprietary structured packing[a] operated at 62.7% of its capacity compared to 79.4% of the existing packing (providing an 16.7% additional capacity). This indicates better column performance with the new packing if the "exponential increase in pressure drop of the existing packing" could be resolved. **TABLE 1** shows capacity and pressure drop values for the existing packing and the new structured packing[a] from hydraulic calculations.

Replacement of existing liquid distributor. Precise liquid distribution is crucial for the performance of packed beds, and distributor selection is vital to achieve maximum packing performance—even the best packing will never provide its full performance if the related internals are not designed appropriately. In a simulation study, the authors' company noticed that the existing liquid distributor drip point density was limited for distribution. The existing liquid distributor was replaced with a high-performance liquid distributor[b] (with a drip point density of 116 holes/m^2) to improve liquid distribution. In turn, this provided better performance in product recovery and delayed plugging problems.

Revamp results. The replacement of the entire packed bed internals and distributor was conducted without any hot work. On startup, the pressure drop across the packing was reduced. The column has been in operation for 8 mos without any operational problems compared to the cleaning/washing cycle required every 3 mos before the revamp.

The project shows how proper cooperation between the authors' companies resulted in successful troubleshooting of a long-prevailing problem. The revamp of the column provided energy savings and increased product yield. MAN recovery has been improved compared to the old packing (**FIG. 4**).

This successful revamp illustrates how unexpected downtime for packing washing—a costly operation for any plant operation in the oil and gas industries—can be avoided. The project achieved numerous benefits with a lucrative payback period of 0.5 mos. HP

NOTES

[a] Sulzer's MellapakPlus™
[b] Sulzer's MellaTech™
[c] AspenTech's Aspen Plus

ACKNOLEDGEMENTS

The case study presented here was performed by Sahara International Petrochemical Company (Sipchem) Saudi Arabia. Sulzer extends its thanks to Ali S. Al-Wadie (Process Engineer and Control System Manager) for his encouragement and engineering approval for this project, and Mohammad M. Al-Zahrani (Operation Manager) for his professionalism, approval and coordination provided within the team during project execution.

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Green refinery challenges: Small-scale sulfur recovery

As the gavel dropped, signaling a historic agreement at the conclusion of COP26 (2021 United Nations Climate Change Conference) in Glasgow, Scotland, leaders from 197 nations agreed to continue supporting the goal of limiting global warming to 1.5°C above pre-industrial levels. The race is on!

Two initiatives that emerged from COP26 were the Mission Innovation and Carbon Dioxide (CO_2) Removal Projects. Mission Innovation—a collaboration among governments to unlock affordable decarbonization pathways—seeks to accelerate technologies that will reduce emissions by the sectors responsible for 52% of current global emissions. The Netherlands and India are leading a biorefinery program to make bio-based alternative fuels and chemicals economically attractive. The other initiative is the Carbon Dioxide Removal Project, which is led by Saudi Arabia, the U.S. and Canada. The project's goal is to net an annual reduction of 100 MMtpy of CO_2 by 2030. Other countries are following suit.

The biggest culprit. In November 2021, the Global Carbon Project, a Global Research Project of Future Earth and a research partner of the World Climate Research Programme, reported that global CO_2 emissions from fossil fuels were 34.8 $GtCO_2$, a decrease of 5.4% from 36.7 $GtCO_2$ in 2019; however, projected global fossil CO_2 emissions in 2021 were forecast to rebound close to their pre-COVID levels after an unprecedented drop in 2020. Emissions from coal and gas use are set to grow more in 2021 than they fell in 2020, but emissions from oil use remain below 2019 levels.

In the U.S. alone, the U.S. Environmental Protection Agency (EPA) reports that the transportation sector generates the largest share of greenhouse gas (GHG) emissions (~ 23%). The Intergovernmental Panel on Climate Change states that this sector presents the most challenges to mitigation.

Energy transition and refinery restructuring. In the context of fluctuating market conditions for traditional oil-derived fuel markets, refiners are increasingly focused on the energy transition to improve the profitability of their assets and secure long-term operations, in tandem with reducing GHG emissions and moving towards carbon neutrality. Many refineries around the world are considering renewable fuels production at either existing refineries that will continue to process crude oil, or at facilities that are idle with existing infrastructure to accommodate new processing units to refine (mainly) diesel and jet fuels from sustainable sources, such as used cooking oils (UCO), waste animal fats (tallow) and/or certified sustainable vegetable oils, such as rapeseed.

With the new processing technologies, there is a co-production of CO_2 and the need to maintain active catalyst in the reaction system. A small amount of sulfur, usually in the form of liquid disulfide oils, is added to the conversion reactors to maintain catalyst activity. The sulfur is converted to hydrogen sulfide (H_2S) in the reactor system. An amine unit can remove the unwanted CO_2 from the process, but this also removes the H_2S. The resulting acid gas stream contains far less sulfur than is practical to remove with a typical refinery sulfur recovery unit (SRU) using Claus technology.

In its continued contribution to improving sustainable mobility, a multinational oil and gas company based in Europe uses a proprietary sulfur recovery technology[a] to serve as the SRU. This technology is a patented, wet scrubbing, liquid redox system that uses a chelated iron solution to convert H_2S to innocuous, elemental sulfur. It is designed to remove about 4 metric tpd (tonnes per day) of sulfur from an acid gas with up to almost 7 mol% H_2S.

The processes. The hydrotreated vegetable oil (HVO) process produces renewable fuels via hydrogenation and hydrocracking of vegetable oils and animal fats using hydrogen and catalysts at high temperatures and pressures. The oxygen is stripped in a reactor with a catalyst that often requires sulfiding to promote conversion chemistry and avoid catalyst deactivation. Dimethyl-disulfude (DMDS), the most commonly used chemical for sulfiding the catalyst, is converted to H_2S, and the oxygen from the feedstock is converted to CO_2 and water during the deoxygenation reaction.

Hydrocarbon liquids exit the reactor and are routed to a three-phase separator, where the water is removed and vapors are collected for cleaning and recycling. The hydrocarbon liquids are routed to an isomerization reactor before being separated into various cuts to create light fuels, sustainable aviation fuels and renewable diesel.

The vapors from the separator are often routed to an amine unit to remove the H_2S and CO_2 formed in the deoxygenation reactor. The treated gas from the amine unit is rich in hydrogen and recycled to the HVO process. A slipstream may be used as fuel gas. Depending on what other units may be operating at the refinery, additional acid gases may be processed in other amine units at the site. The European refinery mentioned here utilized large Claus units that had been previously idled.

FIG. 1. Anadarko Petroleum Corp.'s Pinnacle Bethel Gas Plant, near Bethel, Texas, with the addition of the proprietary sulfide treatment system[a].

SRU process selection. In the European refinery example, the existing higher capacity Claus SRU had been shut down when oil processing ceased. The typical minimum H_2S content for a feed to a Claus unit is 35 mol%, but that condition cannot be met when the HVO process is operating. Modified Claus units can operate with less H_2S in the feed stream, but to achieve >99.9% sulfur recovery, a tail gas treating unit is required downstream of any Claus SRU, which increases cost. Turndown conditions can also negatively affect Claus unit sulfur recovery efficiency.

A highly flexible sulfur recovery process was needed to account for the range of feed gas conditions. The selected process also needed to achieve high recovery percentages over a wide range of operating flowrates and H_2S concentrations. The refiner required:

- Maintaining sulfur removal efficiency at turndown conditions
- No minimum H_2S concentration in the feed acid gas stream
- A fully guaranteed product to meet customer requirements.

The sulfur recovery process[a]. The author's company's process[a] is a liquid redox system that converts H_2S to solid elemental sulfur (**FIG. 1**). The process utilizes an aqueous iron solution with a catalytic performance that is enhanced by a proprietary blend of chemicals. The H_2S is converted to elemental sulfur by redox chemistry according to the following overall reaction (Eq. 1):

$$\text{Direct oxidation reaction: } H_2S + 1/2\ O2 \longrightarrow H_2O + S^\circ \quad (1)$$

The redox reaction is conducted in separate sections of the process, as summarized in Eqs. 2 and 3:

$$\text{Absorber: } H_2S + 2Fe{+}{+}{+} \longrightarrow S^\circ + 2Fe{+}{+} + 2H{+} \quad (2)$$

$$\text{Oxidizer: } ½\ O_2 + H_2O + 2Fe{+}{+} \longrightarrow 2OH{-} + 2Fe{+}{+}{+} \quad (3)$$

The exothermic absorber reaction represents the oxidation of H_2S to elemental sulfur and the accompanying reduction of the ferric iron state (active) to the ferrous iron state (inactive). This reaction is irreversible and not equilibrium dependent. The oxidizer reaction (also exothermic) represents the oxidation of the ferrous iron back to the ferric iron state.

Liquid redox technology produces a sulfur "cake" that contains approximately 65% sulfur, 35% moisture and trace amounts of proprietary chemicals. It does not produce any other products or byproducts, so no additional treatments are required. Because the reaction occurs in the aqueous phase at near-ambient temperatures, the process is inherently safe. The removal of sulfur is not affected by turndown conditions, and the process can be applied to almost any vapor stream containing H_2S.

Liquid redox is an economical process for H_2S removal and sulfur recovery within a given range. On the low end of sulfur recovery, non-regenerative chemical absorbents (scavengers) are likely to be more economical due to the low capital cost of a two-vessel design. As the sulfur removal requirement increases, operating costs become prohibitive using scavengers. This is when a liquid redox process is advantageous due to lower operating costs. The higher capital cost of a liquid redox unit is quickly paid out in operating savings vs. replacement of the non-regenerative scavenger. An amine unit + Claus SRU is advantageous for high-end sulfur recovery. Excessive loss of sulfur recovery[a] chemicals when removing more than 20 tpd of sulfur cake, however, drives up costs.

The first milestone. A proprietary sulfur recovery unit[a] went into commission at the European refinery in August 2019. The refinery installed a rectangular autocirculation unit enabled by the dilute concentration of H_2S in the feed to the SRU. In the summer of 2020, the liquid redox unit underwent its first turnaround. The working solution was drained and stored for further use. The autocirculation vessel was opened, cleaned and inspected. A small amount of sulfur buildup was detected on the vessel's floor. The unit was cleaned with a warm water wash and no corrosion was noted. The shutdown for maintenance took 1 wk—the equivalent to 98% availability in the first year of operation.

In 2020, the biorefinery reached full operation with a five-fold increase in biofuels production compared to 2019.

Takeaway. The rising demand for sustainable and reliable energy is driving biofuel production around the world. Analyst firm Precedence Research predicts the global biofuels market will exceed $200 B by 2030.

Slight reconfigurations of existing petroleum refineries enable a seamless and lower-cost entry into the biofuels production business at scale. Refineries that will continue to process crude oil will typically utilize existing Claus units to handle any increased demand for sulfur recovery. In facilities without a working Claus unit or where the Claus unit is overloaded, bio-refiners can manage the low-tonnage sulfur recovery with the proprietary liquid redox technology[a], which addresses low-tonnage sulfur recovery and achieves very low SO_2 emissions. The systems are reliable and have high availability, the operating cost is low, and bio-refiners can consistently achieve sulfur removal guarantees while taking a solid stance on sustainability. HP

NOTES

[a] Merichem's LO-CAT® process (Liquid Oxidation CATalyst)
[b] Merichem's FIBER FILM®

WILLIAM ECHT has worked in gas conditioning for more than 40 yr. He serves as Technology Licensing Director for Merichem Co., representing the technology[a] for sulfur recovery from gas streams and its technology[b] for sulfur removal from liquid hydrocarbons. After receiving a BS degree in chemical engineering from the University of Texas at Austin, Mr. Echt joined Phillips Petroleum in Oklahoma for 8 yr as Staff Engineer for gas gathering and natural gas liquids (NGL) extraction plants. He then worked for UOP for 25 yr as a Process Engineer and Manager for gas treating technologies based on adsorbents, solvents and membranes.

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Choosing the proper regulator configuration can reduce droop

Many factors go into selecting the proper regulator for your system. One of those factors is sensitivity to changes in outlet pressure as downstream flow fluctuates, which is often characterized as "droop" as illustrated within a regulator's flow curve.

What is droop? Droop is defined as the reduction of outlet pressure experienced by pressure-reducing regulators as the downstream flowrate increases. While every pressure-reducing regulator will exhibit some droop, external components can be added to the regulator configuration to help maintain reliable downstream pressure and minimize droop.

Flow curves are a useful tool in understanding the range of outlet pressures a regulator will maintain based on various system flowrates. They are created through product testing and represent the real performance of a regulator for a given set of system parameters.

FIG. 1 is an example of a flow curve with four different curves shown to demonstrate how different pressure regulation configurations can help to reduce droop. The vertical axis displays outlet pressure, with the horizontal axis showing downstream flowrate. The flatter or more horizontal the curve, the more the regulator preserves consistent pressure. For example, the purple curve—which is a result of a configuration using a dome-loaded regulator and a pilot regulator with an added feedback line to the pilot regulator—shows the best performance. The far right of each curve indicates that the pressure will quickly decline toward zero when the regulator is completely open. Under these circumstances, the poppet is pushed open toward its stroke limit. At this point, the regulator acts more like a restricting orifice than a pressure control device.

No matter how good your regulator is, some droop will be experienced. However, the goal is to flatten the curve as much as possible. Selecting the proper regulator configuration for your system will enable you to achieve this goal.

Option A: A simple spring-loaded regulator. The most common type of pressure-reducing regulator is a spring-loaded regulator (**FIG. 2A**). In this design, a spring applies force on a sensing element—either a diaphragm or a piston—which moves the poppet closer to (**FIG. 2B**) or farther away from the orifice.

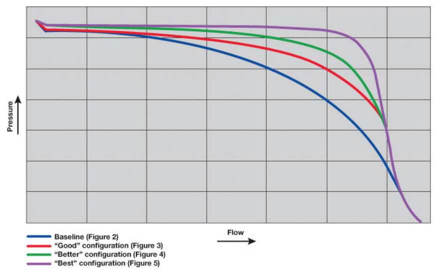

FIG. 1. This chart illustrates different flow curves using different pressure regulation configurations: a simple spring-loaded regulator (Option A—Baseline); a dome-loaded regulator and a pilot regulator (Option B); a dome-loaded regulator and a pilot regulator with an added feedback line to the dome-loaded regulator (Option C); and a dome-loaded regulator and a pilot regulator with an added feedback line to the pilot regulator (Option D).

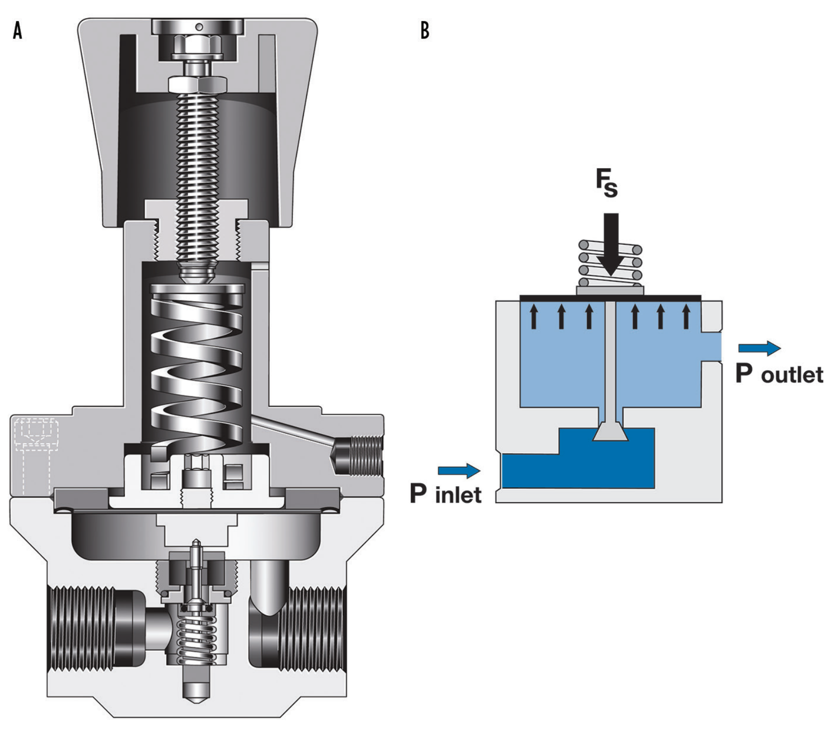

FIG. 2. A spring-loaded regulator (A), which uses a spring to control flow and pressure by moving the poppet either closer to or farther away from the orifice (B), provides a baseline (Option A) for the flow curve comparison.

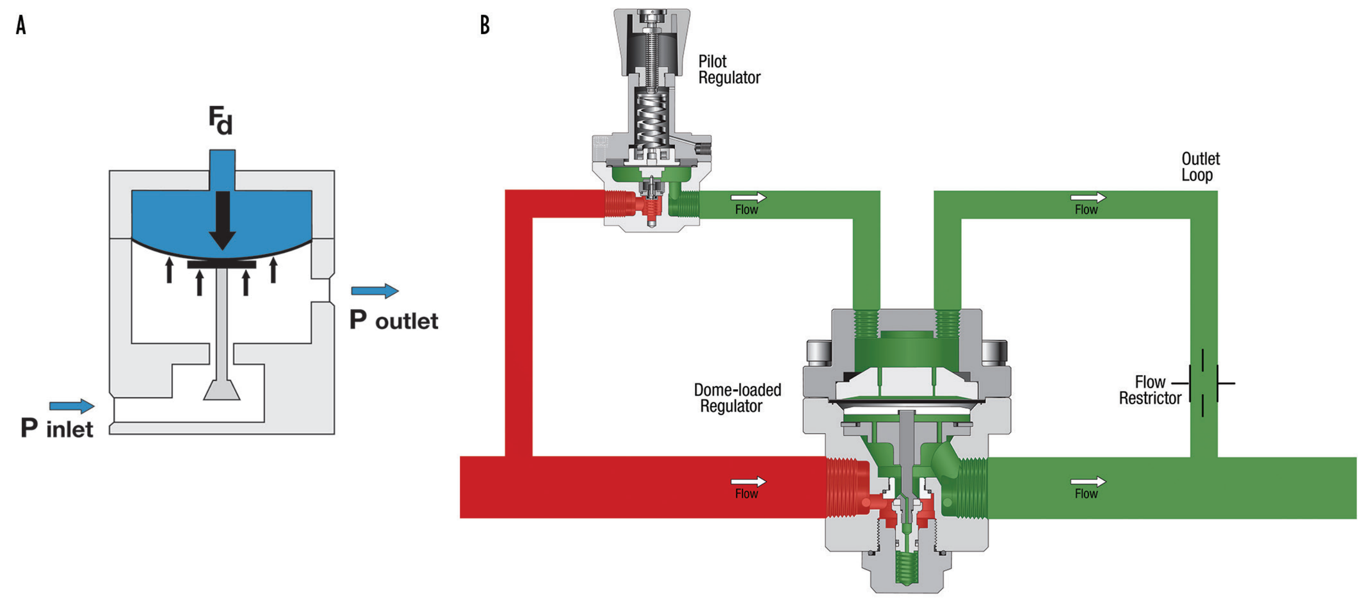

FIG. 3. A dome-loaded regulator uses pressurized gas housed in a dome chamber to flex a diaphragm and move the poppet to control downstream pressure (A). The Option B configuration (B) features a dome-loaded regulator with a pilot regulator and a dynamic control outlet loop to control dome pressure.

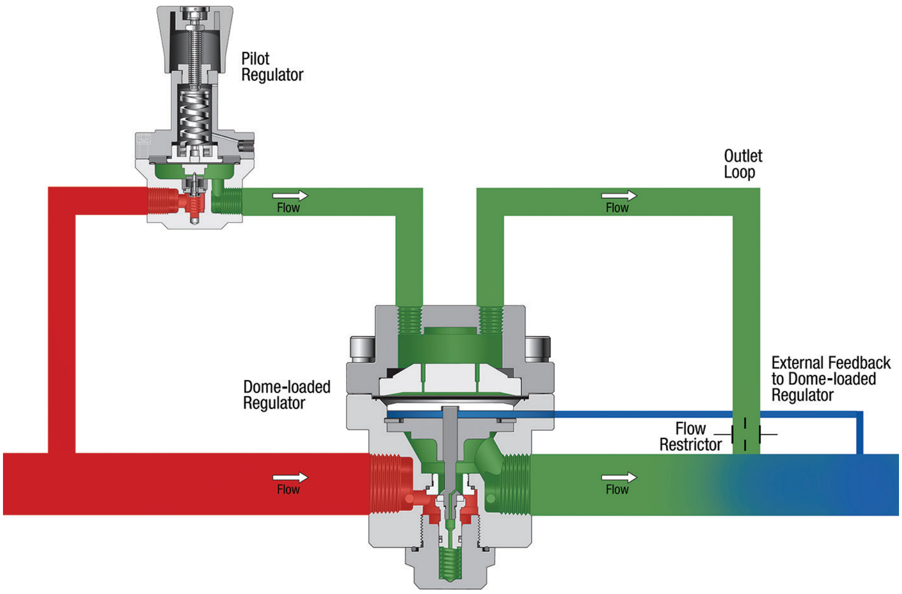

FIG. 4. The Option C configuration features an external feedback line connected to the dome-loaded regulator to better compensate for downstream pressure drops.

This controls the downstream pressure. The spring-loaded regulator is then used for purposes of comparison as the baseline (Option A) in **FIG. 1.**

Generally, a spring-loaded pressure regulator will reduce droop adequately for most basic applications. The regulator poppet will then allow additional flow by moving away from the seat as system flow demand increases. When this happens, the loading spring will relax and lower the loading force and regulator setpoint. The amount of droop this system allows depends on the loading spring rate as flow demands fluctuate. To manage the spring rate, technicians may have to frequently perform manual adjustments to maintain the desired set pressure, particularly if a high degree of accuracy is necessary.

Even though this system can be more effective than a simple spring-loaded regulator in many cases, its droop-reduction performance can be enhanced by additional components and design modifications.

Option B: Dome-loaded regulator with a pilot regulator. As the industrial fluid system becomes more sophisticated, so do the mechanisms to maintain adequate pressure. It may be necessary to employ a dome-loaded, pressure-reducing regulator as the complexity of the system increases. In this system, the loading is managed by a pressurized gas housed in a dome chamber. The gas flexes a diaphragm that moves the poppet away from the orifice and controls the downstream pressure (**FIG. 3A**).

Option B combines a dome-loaded, pressure-reducing regulator with a pilot regulator. This system regulates pressure changes by keeping constant pressure in the dome chamber, while the pilot regulator controls the supply of gas to the chamber. As shown in **FIG. 3B,** any excess dome pressure is relieved through an outlet loop.

As with the spring-loaded regulator, the poppet will allow additional flow as the system flow demands increase, though the mechanism is quite different. In this system, the diaphragm flexes downward to lower the dome pressure. This triggers the pilot regulator to allow additional gas into the dome so consistent pressure can be maintained.

When flow demands decrease downstream, the diaphragm flexes upward, causing the pressure in the dome to rise. Under these circumstances, the excess pressure vents to the downstream side

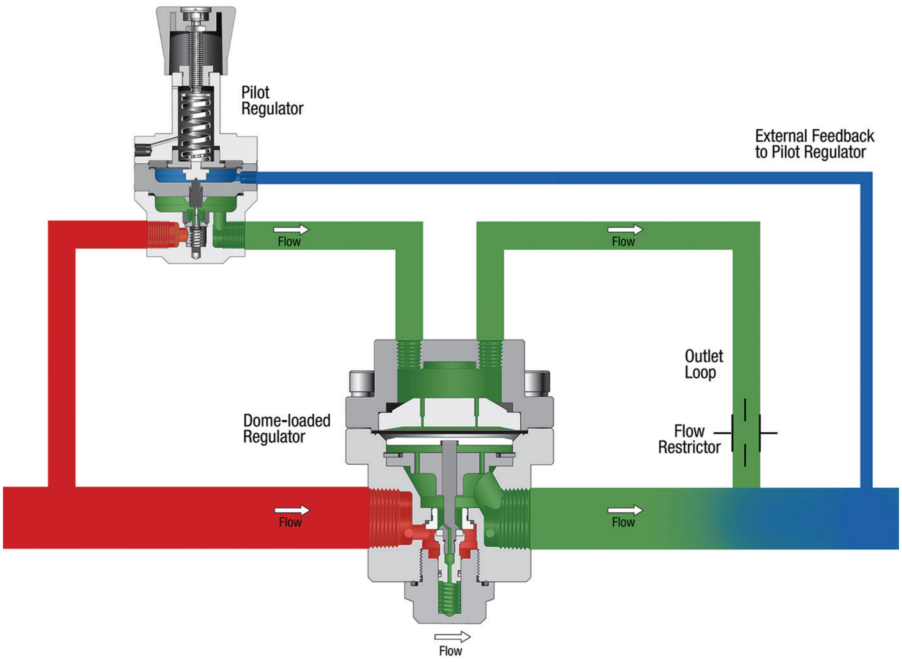

FIG. 5. The Option D configuration features an external feedback line connected to the pilot regulator, delivering downstream pressure feedback.

through the dynamic control outlet loop.

In **FIG. 1,** this system is labeled Option B and shows much more robust pressure control than Option A (the spring-loaded regulator). The amount of droop is reduced, as represented by the flatter flow curve. It is clear that dome regulators equipped with pilot regulators can manage pressure more accurately over a wide range of flows. In most instances, standard dome-loaded regulators are employed with confidence that significant pressure drops will not occur. However, even this system can be improved.

Option C: External feedback line connected to a dome-loaded regulator. Additional accuracy can be achieved by adding external feedback to a dome-loaded regulator, as shown in **FIG. 4.** External feedback is sent to the regulator by connecting a tube from the downstream process line back to the sensing area of the dome-loaded regulator.

The external feedback line provides more precise feedback on pressure levels within the system instead of depending solely on the pressure levels in the regulator itself as is seen in standard dome-loaded regulators. In **FIG. 1,** this system is shown as Option C, where some droop is still observed despite a flatter flow curve than either a spring-loaded regulator or a dome-loaded regulator.

Option D: External feedback line connected to a pilot regulator. In **FIG. 5,** the final option connects the external feedback line directly to the pilot regulator instead of the dome-loaded regulator. This enables the pilot regulator to make highly accurate adjustments to pressure in the dome-loaded regulator's chamber based on actual downstream pressure. Then, the dome-loaded regulator can compensate by changing its outlet pressure.

When system flow demands rise in this configuration, pilot regulators monitor lower pressures with the aid of the external feedback line. If pressures drop, the pilot regulator increases the pressure inside the dome, which maintains the proper downstream set pressure. The feedback loop allows continuous, automatic adjustments that keep the system stable and working at optimal performance. In **FIG. 1,** this is represented by Option D, which clearly has the least droop and the broadest flow range.

Minimizing droop. All regulators will exhibit droop, but the right system configuration can minimize it by keeping the pressure constant as flow changes. Work with your regulator supplier to design the right system for your application. HP

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Evaluate options for decarbonizing petroleum refineries

The energy transition requires rebuilding the energy supply infrastructure for a lower-carbon economy and renewable energy generation for industry and transportation that run more on electricity and hydrogen (H_2) and less on fossil fuels and provide a circular path for consumer plastics. The growth in global oil demand is predicted to end within 10 yr, but it is still too early to foresee a rapid decline in that demand.

Refiners are strategizing to align with a low-carbon future. Moving from transport fuels to petrochemical feedstocks, producing renewable fuels and sourcing low-carbon H_2 are emerging as possible options. However, uncertainty remains around demand forecasts and potential regulatory framework. Governments, technology providers, refinery owners and engineering companies will all play a role in selecting and implementing the most efficient and least-disruptive pathways to a low-carbon future. One thing is certain: this transition will be gradual and require operating refineries to align product slate with demand and reduce their direct carbon dioxide (CO_2) emissions.

Although the vast majority (85%–90%) of carbon emissions in the petroleum fuel cycle occur during final consumption of the petroleum products, refineries remain a substantial source. This article estimates the CO_2 emissions from refineries and the impact that factors such as crude quality, configuration, electricity source, etc., have on these emissions estimates. Options for reducing emissions from major sources are also evaluated.

Refinery CO_2 emissions. Petroleum refining products are an essential part of the world economy, providing energy as well as other high-value products. The refining process involves separating, cracking, restructuring, treating and blending hydrocarbon molecules to generate petroleum products. FIG. 1 shows a simplified, typical overall refining process. The specific processes implemented will vary from refinery to refinery and depend upon the refinery's capacity, feedstock and specific product slate.

Each refinery has unique features, which makes finding a global unified model to represent all of them difficult. This is even more true when it comes to estimating CO_2 emissions from a refinery, allocating the emissions to products to determine the impact on fuel carbon intensity, and developing strategies for CO_2 reduction. However, it is possible to represent operating refineries in a few scenarios by grouping the common variables impacting CO_2 emissions. While this will not have the granularity of an individual refinery model, it does provide useful insights into potential CO_2 reduction strategies, target emissions and achievable reductions. The Petroleum Refinery Life-cycle Evaluation Model (PRELIM) developed at the University of Calgary and freely available for download[1] is used here to estimate CO_2 emissions for a range of refinery configurations, feedstocks, upstream emissions and energy sources. PRELIM was developed to offer a free, flexible, transparent and open-source tool that captures the impact of crude oil quality and refinery configuration on energy use and the environmental impacts associated with refining crude oils.

The estimates from PRELIM are analyzed for CO_2 reduction strategies. Technically feasible technologies are proposed for reducing CO_2 from refinery heating, steam methane reformer (SMR) H_2 units utilizing natural gas (NG) feedstock and fluidized catalytic cracking (FCC). For simplicity, global warming potential (GWP) in terms of CO_2 equivalent (CO_2e) is presented as CO_2—in a refinery, that is the major component.

Modelling methodology. PRELIM employs a system-level approach and refinery linear modelling methods. This

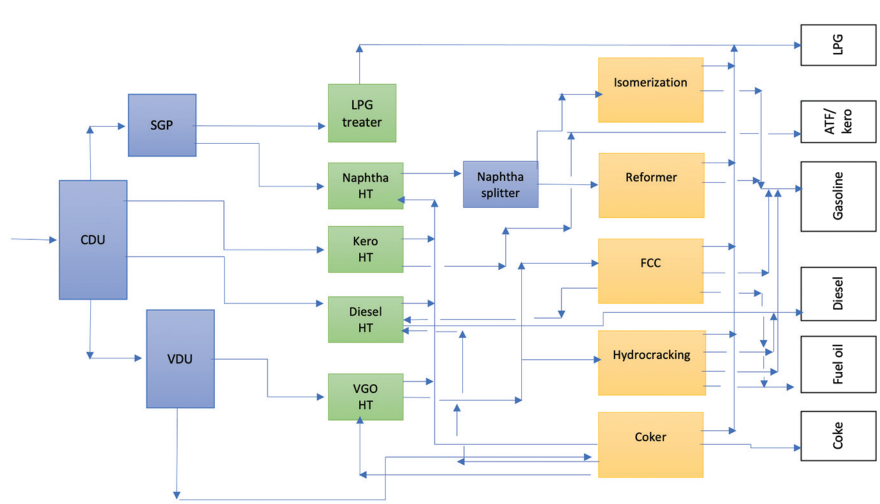

FIG. 1. Simplified flowchart of refining processes and product flows.[2]

allows the selection of different crudes and alternate refinery configurations. It is also capable of changing the product slate, such as the gasoline/diesel ratio and petrochemical feedstocks, alternate dispositions of refinery off-gas, alternate sources of electricity, etc. **TABLE 1** presents the parameters and options used for this article.

Other parameters (e.g., allocation methods, allocation products, GWP values) are kept as default. In this model, a medium-conversion refinery includes crude/vacuum distillation, naphtha hydrotreater, isomerization, naphtha catalytic reformer, diesel/ kerosene hydrotreater, FCC and gasoil hydrocracker units. All medium-conversion units are included in a deep-conversion configuration with the addition of either a delayed coking or residue hydrocracking unit. PRELIM presently does not include configurations with petrochemical integration (e.g., multi-feed cracker, aromatics). The article excludes these integrations and only discusses fuels refining.

The parameters presented in **TABLE 1** were grouped into 11 alternate cases and run in PRELIM. Selected options cover a wide range of existing refineries. **TABLE 2** summarizes the input parameters and es-

TABLE 1. Model parameters and options

Parameter	Options
Crude type	Light sour, medium sour, heavy
Refinery configuration	Medium conversion, deep conversion (coking), deep conversion (hydrocracking)
Naphtha reformer	Straight run, straight run + heavy
SMR H_2 purification	PSA
FCC hydrotreater	Post FCC
Electricity source	Coal-fired, NG-fired, low-carbon
Cogeneration	Natural gas combined-cycle (NGCC), no cogeneration
Off-gas	LPG
Upstream releases	Include, exclude

TABLE 2. Refinery energy use and CO_2 emissions

Parameters	Units	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6	Case 7	Case 8	Case 9	Case 10	Case 11
Crude type		Light sour	Medium sour	Medium sour	Medium sour	Heavy	Heavy	Heavy	Heavy	Heavy	Heavy	Heavy
Crude API	deg API	40	28.5	28.5	28.5	18.3	18.3	18.3	18.3	18.3	18.3	18.3
Crude S%	%	0.88	2.4	2.4	2.4	5.3	5.3	5.3	5.3	5.3	5.3	5.3
Refinery configuration		Medium conversion	Medium conversion	Deep conversion (coking)	Deep conversion (hydro-cracking)	Deep conversion (coking)	Deep conversion (hydro-cracking)	Deep conversion (coking)	Deep conversion (coking)	Deep conversion (hydro-cracking)	Deep conversion (hydro-cracking)	Deep conversion (hydro-cracking)
Upstream releases		Include	Include	Include	Include	Include	Include	Include	Include	Include	Exclude	Include
Electricity source		NGCC	NGCC	NGCC	NGCC	NGCC	NGCC	NGCC	Low-carbon	Low-carbon	NGCC	Coal-fired
Cogeneration		NGCC-CHP	NGCC-CHP	NGCC-CHP	NGCC-CHP	NGCC-CHP	NGCC-CHP	No	No	No	NGCC-CHP	No
Blended gasoline	wt%	40.5	28.8	40.1	37.4	40.6	34	40.6	40.6	34	34	34
Jet-A/AVTUR	wt%	22.4	18.6	19	18.9	6.2	6.1	6.2	6.2	6.1	6.1	6.1
ULSD	wt%	19.8	18,2	24	30.4	25.1	36.5	25.1	25.1	36.5	36.5	36.5
Coke	wt%	0	0	9.8	0	17.5	0	17.5	17.5	0	0	0
Residue hydrocracker Btm	wt%	0	0	0	7.6	0	15.5	0	0	15.5	15.5	15.5
Liquid heavy ends	wt%	15.4	32.6	3.1	2.6	4.1	2.8	4.1	4.1	2.8	2.8	2.8
Sulfur	wt%	0.4	0.8	1.5	1.6	3.1	3.4	3.1	3.1	3.4	3.4	3.4
LPG	wt%	1.4	0.9	2.5	1.5	3.4	1.6	3.4	3.4	1.6	1.6	1.6
Energy use, electricity	MJ/bbl	3.2	2.8	3.6	3.4	5.3	5	26.2	26.2	33.1	5	33.1
Energy use, RFG	MJ/bbl	99.3	74.8	176.4	199.3	258	329	258	258	340.2	329	366.6
Energy use, natural gas	MJ/bbl	241	209.1	225.2	188.8	190.4	104.2	152.9	152.9	42.7	104.2	42.7
Energy use, H_2 via SMR	MJ/bbl	67	83.3	139.4	311.6	221.6	675.9	220.2	220.2	671.7	675.9	671.7
Energy use, H_2 via CNR	MJ/bbl	67.4	41.4	43.5	42.1	13.5	12.8	135	135	12.8	12.8	12.8
Total refinery processes	MJ/bbl	503.1	437.8	626.1	777.5	744.6	1,165.8	726.9	726.9	1,139.6	1,165.8	1,139.6
CO_2, electricity	kg CO_2/bbl	0.5	0.4	0.6	0.5	0.8	0.8	4	0.2	0.3	0	10.9
CO_2, Process heating	kg CO_2/bbl	21.2	17.9	25	24.2	27.9	27.3	23.9	23.9	21.9	26.2	21.9
CO_2, Steam generation	kg CO_2/bbl	1.5	1.1	1.2	0.8	0.7	-0.7	2.1	2.1	1.1	-0.7	1.2
CO_2, H2 SMR	kg CO_2/bbl	4.6	5.8	9.6	21.4	15.2	46.5	15.3	15	45.9	39.2	47.5
CO_2, FCC coke burnofff	kg CO_2/bbl	3	3.1	4.5	3.8	6.6	4.6	6.6	6.6	4.6	4.6	4.6
Total refinery processes	kg CO_2/bbl	31.7	29.3	41.8	51.8	52.8	80.2	53.5	49.4	75.5	710	87.8

timated product slate, energy use and CO_2 emissions. Parameters are varied to bring out the major factors that impact CO_2 emissions. CO_2 emissions are presented as per barrel (bbl) of crude processed to eliminate the impact of refinery capacity. For this article, the default data and constants in PRELIM are used for estimates. PRELIM does give the option to modify these defaults values with data available for specific design.

The input parameters that are varied are: crude quality; refinery configuration; cogeneration, included or not; source of electricity; and upstream emissions for NG, included or excluded. This covers a wide range of existing refineries and captures the variations in CO_2 emissions both at an aggregate level and specific to the source.

Discussion of modelling results. From **TABLE 2,** CO_2 emissions range between 30 kg CO_2/bbl and 90 kg CO_2/bbl of crude processed. Major findings are summarized below:

1. For the same crude (Cases 2–4), moving from a medium-conversion to deep-conversion refinery with a coker increases CO_2 emissions by ~40%. Processing the same crude in deep conversion with a residue hydrocracker further increases emissions by ~20%.
2. For a refinery with the same configuration (a deep conversion with coker), processing heavier crude (Cases 3 and 5) increases CO_2 emissions by ~25%.
3. The difference in CO_2 emissions between a deep-conversion refinery with a residue hydrocracker vs. one with a coker is 50% (Cases 5 and 6).
4. For the same crude and refinery configuration, the decrease in CO_2 emissions by switching from a natural gas combined-cycle (NGCC) power plant, or cogeneration based on NG, to a low-carbon power source is ~8% (Cases 5, 7 and 8).
5. For the same crude and refinery configuration, the decrease in CO_2 emissions in switching from a coal-fired power plant source to a low-carbon power source is ~14% (Cases 9 and 11).
6. Upstream NG emissions can account for ~10% of the total estimated CO_2 emissions (Cases 6 and 10).
7. Process heating accounts for 30%–65% of the total CO_2 emissions. This share is higher for refineries processing lighter crudes.
8. H_2 production from the SMR can account for 15%–60% of the total estimated CO_2 emissions. This share is much higher for refineries with residue hydrocracking.
9. An FCCU accounts for 6%–12% of CO_2 emissions and is the next highest single source after process heating and H_2 production.

These conclusions are based on per/bbl basis and, therefore, crude capacity is not a factor. However, depending on crude quality and refinery configuration, the product slates can differ. An alternate perspective of looking at refinery CO_2 emissions is by allocating the CO_2 to the products. This perspective can then be utilized in estimating the carbon intensity of fuel products. This aspect is not discussed in this paper. PRELIM estimates are based on calculations and equations available in literature, and it allows expert input to correct for design when used for a specific site.

CO_2 REDUCTION OPTIONS

Refinery configuration, crude selection and product slate are determined based on economics. Results presented and discussed in the previous section indicate the possibility of reducing CO_2 emissions by processing a lighter crude. This will be feasible if the additional cost of lighter crude is compensated by being eligible for trading CO_2 saved as a result. Similarly, for new refineries in the planning stage, CO_2 emissions can be included in internal rate of return (IRR) calculations by assigning a price to the emitted CO_2. This approach will penalize configurations with higher CO_2 emissions. These options are not further elaborated here since the focus is on existing refineries and on deeper reductions than offered by crude switching or change in configuration.

Another option for refineries is to source electricity from a low-carbon source. For refineries operating with electricity sourced from coal-fired plants, this provides substantial (~14%) CO_2 reductions. For refineries operating with cogeneration utilizing NGCC, the benefits are lower (~8%). This option requires a review of the overall refinery steam and power balance and investigating the conversion of some of the steam-driven equipment to electric drives to decrease CO_2.

To achieve deeper reduction in CO_2 emissions, refineries must reduce emissions from three sources: process heating, H_2 production and FCC catalyst regeneration. Three main approaches to reduce CO_2 from these areas are pre-combustion, post-combustion and oxy-fuel combustion.

Post-combustion systems separate CO_2 from the flue gases produced by the combustion of the primary fuel with air. Pre-combustion systems process the primary fuel in a reactor with steam and air or oxygen to produce a mixture of carbon monoxide (CO), CO_2 and H_2 (synthesis gas). Additional H_2, together with CO_2, is produced by reacting the CO with steam in a second reactor (shift reactor). The resulting mixture of H_2 and CO_2 can then be separated into a CO_2- and a H_2-rich stream. The CO_2 is captured, and the syngas or carbon-free H_2 can be combusted to generate power and/or heat, or utilized for synthesizing chemicals.

Oxy-fuel combustion systems use oxygen rather than air for combustion of the primary fuel to produce a flue gas that is mainly water vapor and CO_2. This results in a flue gas with high CO_2 concentrations (> 80 vol%). The water vapor is then removed by cooling and compressing the gas stream. While all three are technically feasible, considering the need for an air separation unit, extensive oxygen piping, retrofit of flue gas recycle, sealing of furnaces, etc., the option of oxy-fuel firing is not considered in this article. The other two options are discussed in detail in the subsequent sections for each type of emissions source.

CO_2 reduction from refinery heaters. CO_2 emissions attributed to process heaters are spatially distributed across the refinery. **TABLE 3** presents typical source-wise emissions for a deep-conversion coking refinery processing 100,000 bpd.

Flue gases from these furnaces have high volumetric flowrates, are at atmospheric pressure and have low CO_2 partial pressure. Despite the low CO_2 partial pressure, certain amines—such as monoethanolamine (MEA) and other

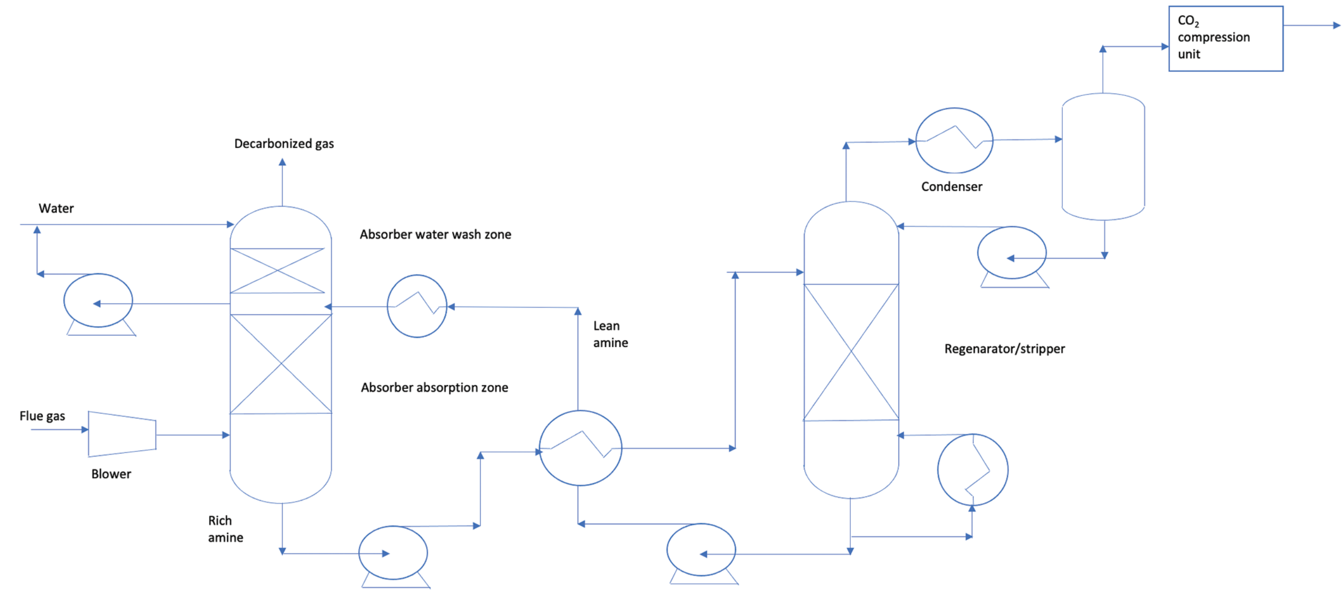

FIG. 2. CO_2 capture utilizing amine or similar chemical solvents for low-pressure streams.

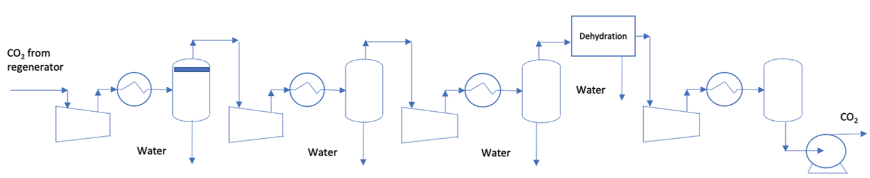

FIG. 3. CO_2 compression unit.

TABLE 3. Estimated CO_2 emissions from furnaces

Furnace	Duty, MW	CO_2, tpd
Atmospheric tower	73	370
Vacuum tower	32	160
Naphtha hydrotreater	11	60
Catalytic naphtha reformer	22	110
Kerosene hydrotreater	12	60
Diesel hydrotreater	18	90
Gasoil hydrocracker	31	160
FCC post hydrotreater	20	100
Coker furnace	97	500
Coker naphtha hydrotreater	23	120

similar solvents—can achieve high levels of CO_2 capture due to fast kinetics and strong chemical reactions. Post-combustion CO_2 capture requires ducting, fans, cooling of flue gas, a CO_2 absorber, a regenerator and CO_2 compression/drying (**FIGS. 2** and **3**).

Low pressure requires large ducts and fans, which consume significant power; to be economical, the CO_2 capture equipment must be located near the emissions source. This imposes limitations for retrofit applications that typically do not have space for ducting, large cooling equipment and absorbers. Additionally, to lower the capital cost, the CO_2 capture units must exploit economies of scale. Providing a dedicated CO_2 capture facility for each furnace in the above refinery example is not an economic choice. The present state of the art design for CO_2 capture from low-pressure streams can handle ~3,000 tpd in a single absorber train. Therefore, designing for CO_2 recovery of ~500 tpd will not achieve economies of scale.

Pre-combustion CO_2 capture applied to refinery process heating requires the conversion of the refinery fuel to decarbonized H_2, as discussed in the previous section. Because of the centralized production of fuel for the entire refinery, pre-combustion can be built at economic scale capacities. The next section discusses CO_2 capture from an SMR H_2 production unit. Many existing refineries have SMRUs to meet process H_2 needs. A dedicated SMR H_2 unit with carbon capture is required to produce H_2 for process heating. The technical feasibility of using H_2 in refinery heaters has been ascertained.[3]

CO_2 reduction from SMR H_2 production. An SMR plant consists of four major sections:

- Feedstock pre-treatment
- Steam reformer
- Shift reactor
- Pressure swing adsorption (PSA) unit.

CO_2 is generated by the reforming and water-gas shift reaction, the combustion of CO in the PSA tail gas, and the combustion of natural gas as supplementary fuel. This implies that CO_2 could be captured from three possible locations (**FIG. 4**):

1. Shifted syngas
2. PSA tail gas
3. SMR flue gas.

TABLE 4 presents the typical pressure, CO_2 mol% and CO_2 quantity in

these three streams for a 70,000-NM^3/hr H_2 plant meeting the requirement of a 100,000-bpd deep-conversion refinery with a coker.

The stream at the inlet of the PSA unit has high operating pressure and high partial pressure of CO_2. CO_2 capture from streams with high partial pressure of CO_2 is proven using methyldiethylamine (MDEA). Due to the high pressure, a fan is not required, and piping and absorber sizes are smaller compared to low-pressure streams. MDEA is a tertiary amine with high CO_2 absorption capacity, low regeneration energy compared to primary amine [such as monoethanolamine (MEA)], but also has a slower reaction rate.

Piperazine is typically added as an activator to MDEA to increase the rate of reaction. Typically, a recovery of 90% is achievable. In the above reference plant, this would give 250,000 tpy of CO_2 recovery. For a deep-conversion refinery with a residue hydrocracker, applying CO_2 capture on this stream alone can yield 750,000 tpy of CO_2. This scheme is similar to that depicted in **FIG. 2,** except that a fan is not required.

Recovery from this high-pressure stream alone yields 52%–55% of the total CO_2 from an SMRU. The PSA outlet stream is at a lower operating pressure but higher CO_2 content. A compressor will be required to boost pressure to approximately 10 barg to use the MDEA process. Equipment sizes required will increase and additional energy will be needed for compression. Design pressures will reduce compared to Option 1. A preliminary analysis indicates this option will be uneconomical at most sites and is not discussed further here.

Capturing CO_2 from reformer flue gases can lead to much higher CO_2 recovery. Typically for these low-pressure streams, primary amines such as MEA or similar solvents are used. These have limited absorption capacities and usable solution concentration ranges. The energy required for regeneration of solvent is significantly higher compared to MDEA. Low pressure implies large ducts and absorber sizes. In the above reference plant, this option would capture ~500,000 tpy of CO_2 recovery.

CO_2 absorption is most efficient at gas temperatures between 30°C and 50°C. This implies that when capturing CO_2 from flue gas, there will be a cooling requirement before routing to an absorber. This adds additional equipment and the need for either cooling water or additional energy to use air cooling/a gas-gas exchanger, significantly impacting the economics of CO_2 recovery.

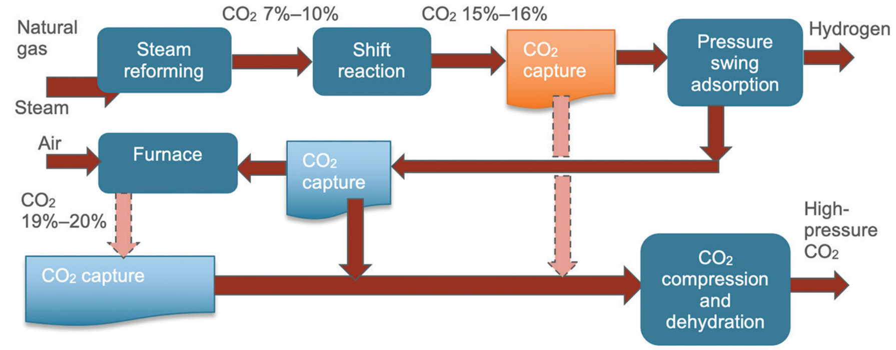

FIG. 4. H_2 production via SMR with three carbon capture options.

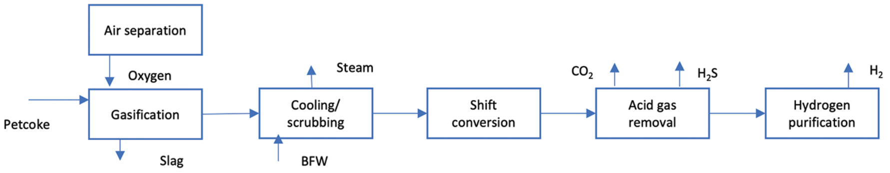

FIG. 5. H_2 production from gasification.

TABLE 4. CO_2 streams in SMR H_2

Stream	Pressure, barg	Temperature, °C	CO_2, mol%	CO_2, tpd
PSA inlet	26	35	16	870
PSA tail gas	1.3	30	50	870
Flue gas	0	130	20	1,450

The above options are applicable for refineries with existing SMR-based H_2 production units. Refineries also have an option to explore alternate routes for meeting H_2 requirements. This may include:

- Green H_2, which is produced by electrolyzing water with the use of renewable energy
- H_2 production from alternate fossil feedstocks, such as petroleum coke/coal gasification with carbon capture.

Green H_2 is a promising alternative in areas with abundant and uninterrupted renewable energy and will become more attractive as capital costs reduce in the future. This option is not discussed further in this article. The option of integrating petroleum coke/coal gasification with CO_2 capture is another promising alternative for low-carbon H_2, and is briefly discussed in the next section.

Gasification for H_2 production.

Gasification is the partial oxidation of any fossil fuel to synthesis gas (syngas), in which the major components are H_2 and CO. To produce H_2, syngas is routed to the shift unit, where CO is shifted to H_2 and CO_2. An acid gas treatment unit removes hydrogen sulfide (H_2S) and CO_2. Shifted and cleaned syngas is routed to the PSA unit where H_2 is recovered, and the rejected tail gas is available for use as fuel (**FIG. 5**). A two-stage shift increases the H_2 content in the fuel and maximizes the degree of CO_2 removal. Gasification requires pure oxygen, which is produced in the air separation unit (ASU).

Feedstock for gasification can be petroleum coke, coal or heavy liquid ends from refining. High-pressure steam is co-produced from heat recovery in syngas and PSA tail gas as fuel. The gasification system and refinery operations can share amine stripper, sulfur block, water treatment and

TABLE 5. Evaluation of technology options for refinery decarbonization

Parameters/Technology	Capital cost	Operating cost	Commercial history	CO_2 abatement	Constructability
CCS for furnace(s)	Strongly unfavorable	Strongly unfavorable	Unfavorable	Strongly favorable	Strongly unfavorable
Hydrogen fuel for furnace, SMR	Neutral	Neutral	Favorable	Strongly favorable	Strongly favorable
CCS upstream of PSA SMR	Favorable	Favorable	Favorable	Neutral	Favorable
CCS from flue gas of SMR	Neutral	Unfavorable	Unfavorable	Favorable	Neutral
CCS FCC	Unfavorable	Unfavorable	Unfavorable	Neutral	Unfavorable
Gasification for H_2	Unfavorable	Unfavorable	Strongly favorable	Favorable	Favorable
Legend	Strongly favorable	Favorable	Neutral	Unfavorable	Strongly unfavorable

cooling water systems. Product compositions vary depending upon the selected gasification technology and the characteristics of the petroleum coke. Coke produced from the 100,000-bpd deep-conversion coking refinery processing heavy crude discussed in previous sections will yield approximately 130 MMsft³d of H_2. This will exceed what is required by the refinery for process use alone. Additional available H_2 can be mixed with refinery fuel gas and used in furnaces, leading to significant additional CO_2 reduction.[3,4]

Since petroleum coke has a significantly higher carbon-to-H_2 ratio compared to NG, the CO_2 quantity that must be captured and sequestered in this route will be significantly higher than in the SMR H_2 route. The electricity requirement will also be significantly higher. However, in this route, all CO_2 is captured from high-pressure syngas. This enables a much more efficient CO_2 capture, yielding an overall CO_2 capture efficiency of ~90% while targeting only high-pressure streams. The gasification route has higher indirect emissions due to electricity, whereas the SMR route has higher upstream emissions, depending on the source of NG. If the electricity for gasification is acquired from a low-carbon source, this route becomes attractive. Since most existing refineries have NG-based SMR H_2 units, the capital costs for this option will be high.

CO_2 reduction from fluid catalytic cracking (FCC).

In the FCCU, during the reaction step, coke is formed and deposited on the surface of the catalyst. To regenerate catalyst, the coke is burned in the regenerator with air. This generates CO_2 and is a substantial single-point source of emissions from the refinery. FCC flue gas contains about 10 mol%–20 mol% of CO_2 when running in full combustion mode.

For the 100,000-bpd refinery discussed in previous sections, the estimated CO_2 emissions are 630 tpd (approximately 200,000 tpy) for a 25,000-bpd FCCU. Two options for CO_2 capture from FCCUs are post-combustion technologies: CO_2 absorption and oxy-combustion. Oxy-combustion has been demonstrated in trials[5] as competitive with post-combustion, providing more flexibility and requiring a smaller plot area. Scale-up and commercial demonstration will be required before it is widely adopted.

Post-combustion capture will be similar to what was discussed for flue gas from heaters and SMRUs: a low-pressure, low-CO_2 partial pressure gas requiring MEA or similar solvents is used. A major difference is the presence of contaminants, such as particulates, sulfur oxides (SO_x) and nitrogen oxides (NO_x), which must be brought down to acceptable levels for amine absorption solvent. This will require a pre-treatment step typically utilizing $DeNO_x$ and a wet gas scrubber. The scrubber reduces particulates, SO_x and temperature to acceptable levels. A single train of blower, scrubber, absorption, regeneration and compression will be capable of handling FCCUs processing 100,000 bpd–120,000 bpd.

Evaluation of options.

A typical qualitative comparison of CO_2 reduction options based on capital and operating costs, commercial history, CO_2 reduction potential and constructability are presented in **TABLE 5.** The best-suited option will strongly depend on site-specific factors, such as plot plan constraints, availability of steam and energy cost, source of crude/NG, cost of capital, availability of water, labor cost, etc.

Takeaway.

In the coming decades, petroleum refineries will be required to continue producing transport fuels while producing fuels from renewable feedstocks and diverting some of the lighter ends to petrochemical production. In addition to adapting the product slate to low-carbon needs, refineries will also be required to reduce their own carbon footprints. This article has identified options available to refineries for the decarbonization of production process. Proven and demonstrated technical options exist for achieving a high level of decarbonization from CO_2 emissions associated with process heating and H_2 production. For FCC CO_2, the solution is technically feasible although other options are also under development. The challenge will be to reduce the cost of applying these options and devising regulatory mechanisms to share the costs of decarbonization. These aspects will be discussed in a subsequent article. HP

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Case study: SCR catalyst plugging

Hydrogen (H_2) reforming is a process to produce H_2 from a compound composed of H_2 atoms. A convection pass is used to heat the process during reforming, which creates nitrogen oxides (NO_x). The NO_x-rich flue gas passes through a selective catalytic reduction (SCR) reactor before being exhausted into the atmosphere. Lint-like fibers from the convection pass and refractory liners create an obstruction that builds up and fouls the face of the SCR catalyst (**FIG. 1**).

The result is decreased NO_x reduction. The higher pressure drops across the catalyst and ultimately reduces H_2 output. Typically, this buildup will require an unscheduled outage for manual cleaning, which can take as long as 5 d at an estimated cost of $1 MM/d.

Lab study. In a study, samples of ash collected from the catalyst were sent to a lab for testing. Microscopy images indicated hair-like particles as well as granular particles. These particles combine and create a dryer lint-like fibrous material.

Testing several online cleaning systems, such as sonic horns, produced no discernible results in breaking up the media. However, the use of air cannon blasts sufficiently broke apart the mass. Various styles of screen material and design were then tested to determine the optimal pitch and diameter for operation in reformers.

Mitigation method. The exact cause of the material being caught in the flue gas is unknown, and it is difficult to pinpoint and address all locations where the refractory fiber may break loose. However, once the material breaks free and accumulates on the SCR catalyst, the dP will increase and NO_x removal will decrease.

Bringing the unit offline to vacuum the material is extremely costly and not ideal; therefore, units often run at a reduced output until the next turnaround. The author's company has installed nearly 100 proprietary screens[a] in front of SCRs since the early 2000s, allowing companies to maintain better output.

Case study: Methane reformer in Texas. Based on these successes, the author's company installed its screens[a] upstream of the SCR on a methane reformer in Texas in 2016. The screen is designed to capture these particles upstream of the SCR catalyst.

The screen pitch is smaller than the pitch of the catalyst, as it is critical to catch all particles too large to pass through the catalyst. Therefore, the company selected a screen system designed to achieve sufficient filtration and flow optimization. The pressure drop was also considered during the design phase in selecting the screen pitch. The goal was to install a screen to prevent buildup and minimize pressure drop increase.

Air cannons. Air cannons are required because the particles tend to become trapped on the screen and block the gas flow, much like fibers in a dryer lint trap. When this happens, the gas will often bypass certain catalyst areas and increase NO_x emissions. Air cannons address this challenge in two ways:

- Blasting the particles and breaking them apart so they are small enough to pass through the filter and the catalyst.
- The blast is aimed to push the remaining particles down to the lower collection area where they will not impact performance.

A CFD analysis was performed with and without the screen in place. The results showed that the screen improved the uniformity index from 90% to 92%. The protective screen is also a flow aid—the better the flow, the better the SCR performance (**FIG. 2**).

FIG. 1. Typical buildup and fouling of a reformer SCR catalyst face.

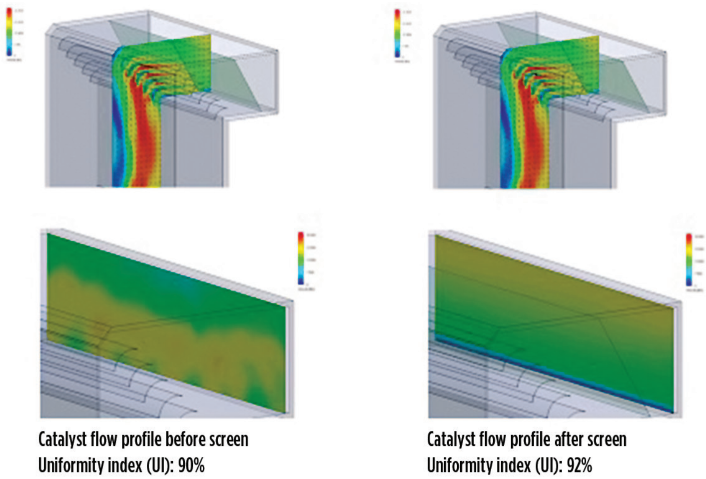

FIG. 2. Catalyst flow profile before and after screen installation.

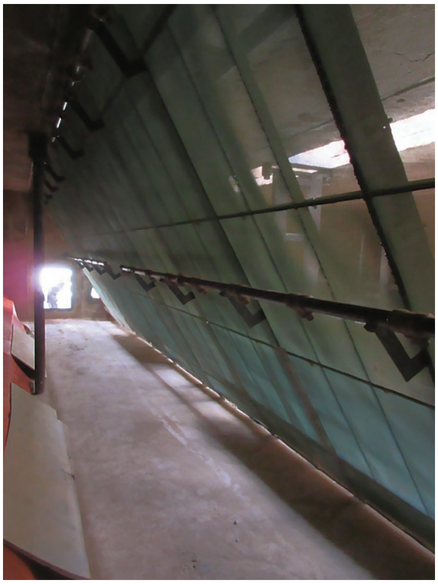

FIG. 3. Completed installation of the filtration screens in a methane reformer.

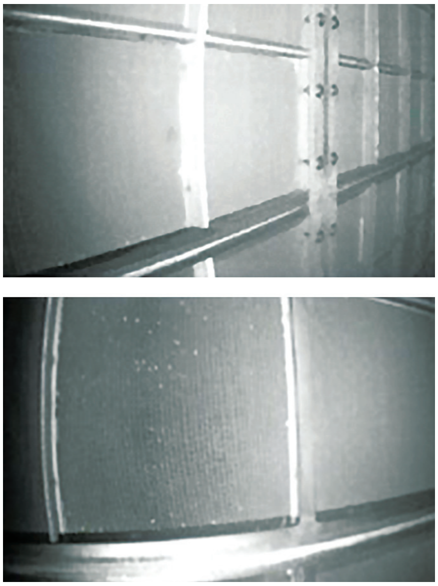

FIG. 4. Catalyst face taken while the unit is online.

Turnaround impact. Due to the standardized modular design, the author's company's mechanical division can install SCR screens, air cannons and access doors in less than a working week of critical path time. In this referenced case, all the work was performed within the time window of the preplanned outage and caused no impact to the turnaround schedule (**FIG. 3**).

Prior to the screen installation, the typical startup sequence required utilizing a cheese cloth to capture particles that break free during startup. The new inline screen system eliminated the need for the cheese cloth by capturing all particles that were too large to pass through the catalyst. This practice was critical before the screen installation because the work performed during the outage resulted in dislodging particles larger than the catalyst opening. This process delayed complete startup by roughly a day. Because of the installation of the screen, this cheese cloth startup process can be eliminated, saving approximately $1 MM during each outage.

Operating assessment. Initially, the Texas plant did not experience any significant pressure drop increase across the catalyst, which was a significant improvement. Before the screen installation, the catalyst required frequent manual cleaning at a significant cost. Initially, the plant opted not to utilize the air cannon system, which eventually caused buildup across the screen. This was predictable, as the screen was designed to capture the material and the air cannons were intended to sweep the buildup off the screen face and into the collection zone. However, even with the buildup on portions of the screen face, the distance between the screen and the catalyst was great enough so that the gas distribution recovered, resulting in significantly improved SCR performance.

Since the air cannons were misused, the buildup on the screen face eventually did reach the catalyst face. Once discovered, the air cannons on the screen were cycled and effectively broke apart the particles on the screen, allowing them to pass through the SCR without causing additional buildup. The particles were visibly seen coming out of the stack, and no pressure drop increase was experienced across the SCR, confirming the successful air cannon cleaning.

The air cannons were not operated regularly to eliminate the risk of material passing through the screen. After 2 yr, it was decided to install a finer pitch secondary screen. The secondary screen eliminated all particle buildup on the catalyst.

Results. **FIG. 4** shows the catalyst face while the unit was online after more than 1 yr of run time, and indicates almost no buildup on the SCR itself; the NO_x reduction numbers remain constant. Due to the infrequent firing of the air cannons, the secondary screen does experience some minor accumulation. However, data indicate this does not impact NO_x removal or SCR performance.

This elimination of buildup and pressure drop eliminated the need for the plant to run at a reduced output, the ID fan is running per design, NO_x reduction of the process gas is optimal and the catalyst life has increased. After recognizing this success, the plant has purchased a second screen system for its sister plant. HP

NOTES

[a] IGS' NoNOx Reformer Screens

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Hydroprocessing catalyst reload and restart best practices—Part 2

Hydroprocessing (hydrotreating and hydrocracking) units are high-pressure, high-temperature units that have multiple reactors, multiple beds per reactor and specialized metallurgy. Catalysts in these units are replaced on a 2 yr–5 yr cycle, depending on feed quality, unit design, catalyst selection, operational constraints and performance.

One of the key drivers for a catalyst change-out turnaround on these high-margin units is minimizing the time that the unit is offline—the cost of downtime is high, particularly when a turnaround is extended.

Managing a catalyst change-out turnaround is vastly different than normal refinery maintenance activities: the complexity is higher, the risks and consequences of unplanned events are much greater, and the required resources are significant but limited. The experience of the key personnel involved is a significant factor during this activity; however, decades of experience are disappearing from refineries due to economic pressures and the age profile of operations/engineering/maintenance personnel.

The responsibilities of refinery process engineers change every 2 yr, so building and promoting experience and expertise in-house for hydroprocessing catalyst reloads is difficult. In the past, the catalyst vendor often provided much of the necessary technical expertise, specifically for the catalyst loading and unit restart activities. However, the impact of the COVID-19 pandemic and ongoing restrictions means that catalyst vendors are generally unable to provide this level of onsite support.

This situation leads to the question of how to best manage the complexities and risks associated with a catalyst change-out and restart for refinery hydroprocessing units, as well as hydroprocessing units for renewable fuels (e.g., vegetable oils and fatty acids).

This article (Part 2 of 2) will discuss the activities associated with catalyst loading and the restart. Part 1 of this article, published in the July issue, discussed the activities associated with turnaround planning and shutdown, catalyst unloading and reactor inspection. The two articles cover the entire process across the full catalyst cycle and many of the best practices used to manage and mitigate the underlying risks for these units.

Catalyst loading. Catalyst is manufactured in the metal-oxide form, but it can also be provided in a presulfided or presulfurized form. **Note:** See the Startup section for an explanation of presulfided and presulfurized catalyst. The two latter forms are more expensive; however, they provide the advantage of reducing the time and complexity of the startup. These catalysts are loaded in an inert (nitrogen) atmosphere—which adds time and complexity to the catalyst loading—although they can be loaded in air, depending on temperature and catalyst storage conditions. An economic evaluation must be made to determine the benefits and risks of using presulfided or presulfurized catalysts over the conventional metal-oxide catalysts in each case.

Catalyst loading is performed using one of two techniques: sock loading or dense loading (shown in FIGS. 5 and 6, respectively). Sock loading is simpler and faster but carries a higher risk of non-uniform loading, leading to a higher probability of flow maldistribution during operation and a shorter run length.

A dense-loading machine spreads the catalyst evenly over the full cross-sectional area of the reactor, normally using shaped ports on a specially designed, variable-speed rotor so that it falls like a uniform rain of catalyst particles. This results in greater alignment of catalyst particles and provides a harder and more homogenous catalyst bed, more evenly distributed flow and a higher catalyst density. The result is 15%–25% higher density for extrudate catalysts and, therefore, a longer cycle, as the run length is proportional to the mass of catalyst loaded, in most cases. Consequently, dense loading is maximized unless pressure drop limitations are reached.

The first bed in a hydrotreating reactor is normally sock loaded, as it usually includes bulk physical filtering materials and/or a graded bed (layers of successively smaller catalyst) to limit pressure drop issues resulting from fine scale and other contaminants in the feed. The demetalization catalysts are also normally sock loaded. The hydrotreating catalyst in the first bed can be sock or dense loaded, or a combination, depending on pressure drop constraints. The bottom portion of the bed will be dense loaded if a combined sock-/dense-loading approach is proposed.

Numerous different proprietary dense-loading machines are used, most of which can provide excellent dense-loading results. It is important to ensure that the technician performing the dense loading is experienced with the machine being used, as the quality of loading will have a direct impact on refining margins for the unit over its subsequent catalyst cycle. Using the lowest cost option for dense loading will likely prove very costly.

Stab-in thermocouple assemblies are withdrawn for dense loading of the catalyst and replaced as the catalyst layer ap-

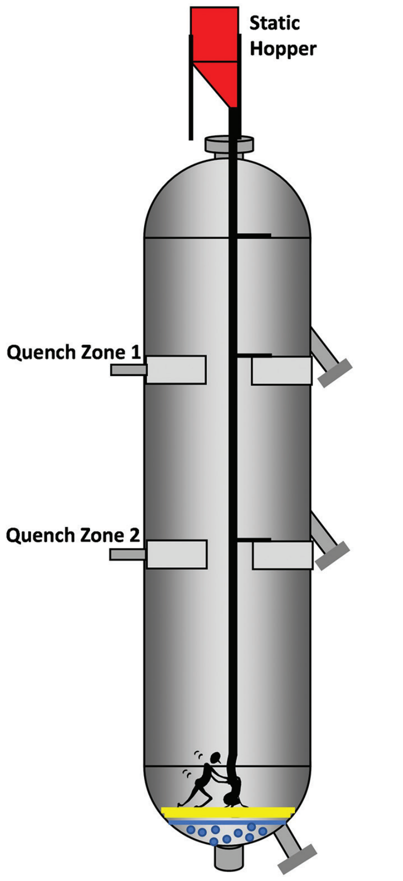


FIG. 5. Sock loading.

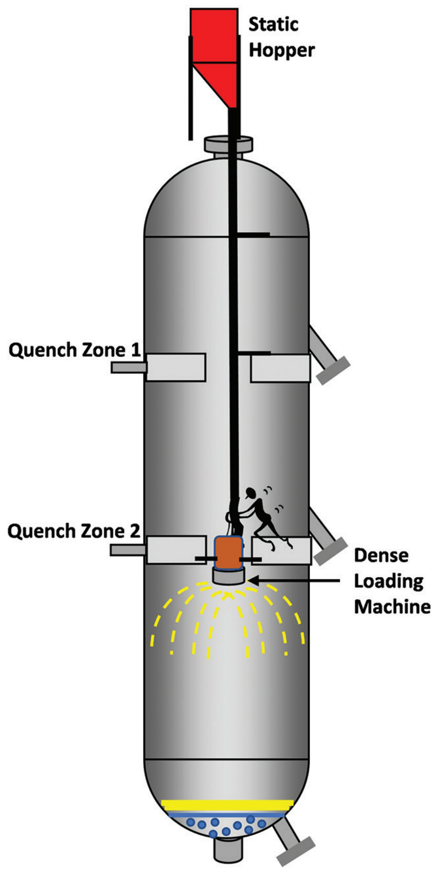


FIG. 6. Dense loading.

FIG. 7. Static (A) and mobile (B) hoppers.

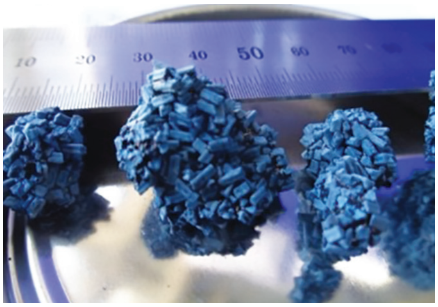


FIG. 8. Agglomerated catalyst.

proaches the height of the nozzle. This is to avoid shadow effects in the dense-loaded catalyst bed.

Before loading begins, a final inspection of the reactor is required to ensure that the reactor, the distribution and other quench zone trays, and the catalyst support grids are clean and free of any fouling material that will affect liquid and gas distribution across the reactor. The catalyst dump nozzles are closed after the ceramic plug and locking plate are installed.

The bed depth of the layers of ceramics and catalyst is normally controlled by a combination of markings on the reactor wall and outage measurements to the ceramic/catalyst bed surface, after ensuring the bed is level. It is vital to confirm that the minimum bed depths for ceramic balls are adhered to, and a method to determine if these layers are uniform (i.e., a flat profile) should be used. This becomes more important as the inside diameter of the reactor increases. This guards against catalyst migration and ensures good liquid distribution during operation.

Equal preparation (or more) is required on the ground as in the reactor(s) for catalyst loading. Cranes must be located so that each lift is simplified to ensure that the catalyst supply to the top of the reactor can keep up with the loading rate of the dense-loading machine. Any unplanned interruptions to dense loading will affect the final loaded catalyst density.

Catalyst is delivered in catalyst bins, bulk bags or drums. Ceramic balls and catalyst should be brought to the site from the warehouse in catalyst bed lots, and catalyst bins/bags/drums must be clearly marked with the catalyst name and size to limit potential mistakes by loading the wrong catalyst in the wrong bed.

A static hopper is installed above the top reactor manway (**FIG. 7A**) to facilitate catalyst delivery into the reactor. Catalyst bins and bulk bags are the most efficient means of storage and handling for catalyst and can be lifted to the top of the reactor to supply catalyst to the static hopper and, from there, into the reactor. Catalyst and ceramic balls delivered in drums or small bags must be decanted into a mobile hopper (**FIG. 7B**) for transporting between the ground and the top of the reactor. All hoppers should have coarse mesh screens installed to avoid ingress of any extraneous material or clumps of agglomerated catalyst (**FIG. 8**) that may block the loading pipe and dense-loading machine. Setting up the static hopper, installing the loading pipe, adjusting the scaffold platforms and weather protection, and finally setting up the dense-loading machine can take several hours.

When loading ceramic balls into the bottom of the reactor, avoid broken or cracked balls in the region around and on top of the outlet collector—the use of spiral inserts, like those shown in **FIG. 9**, in the top of the loading pipe or the use of a proprietary loading system[a], for example, is recommended.

Weather protection will be required in most locations. Most catalysts, especially hydrotreating catalysts, are sensitive to moisture and must be protected from any rain. The hoppers should have fitted lids and the whole area around the reactor manhole and the static hopper should have weather protection. If a drum decanting platform is being used on the ground, it should have its own weather protection, the same for all catalysts stored onsite (bulk bags, drums or bins).

The fresh catalyst sampling protocol must be clearly communicated to the catalyst handling contractor and sample

FIG. 9. Spiral inserts.

containers provided. The catalyst loading must have a robust quality assurance/quality control (QA/QC) process that should include duplicate identification and the counting of catalyst bins/bags/drums loaded into the designated reactor/bed. Each bin/bag/drum number, as well as the catalyst batch number, should be recorded against the respective reactor/bed. This prevents any errors and provides important information if a catalyst performance issue occurs after startup.

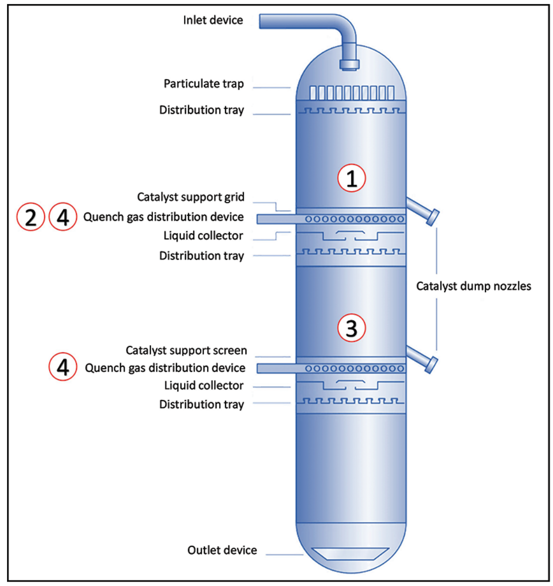
The dense-loading technician will stop loading at intervals to check the bed profile and the catalyst density to ensure the bed is loading consistently flat and not dished or mounded. A dished or mounded profile will affect liquid distribution through the bed during operation. If the bed must be levelled, it is important that a uniform layer of the bed is disturbed and levelled, resulting in a uniform layer of the equivalent of sock-loaded catalyst. Just pushing the mounded catalyst out or filling the dished section into the middle of the bed will cause distribution issues.

The dense-loading machine can only load effectively up to 300 mm–500 mm from the bottom of the distribution tray. The final part of the bed must be sock loaded.

The catalyst bed must be levelled and confirmed as level at each catalyst layer transition and at the top of the bed. This should be verified by physical inspection or, at the very least, a video inspection.

When the catalyst bed is accepted as completed, the distribution tray and other quench zone trays are cleaned again and closed, along with the catalyst support grid for the bed above. The closing of each tray should be witnessed. The process of loading the next bed is then repeated.

Physical filtration materials may be used in the very top of the lead reactor, per stage (i.e., the last layer loaded). This material is often awkward to load and spread in the confined space between the top of a catalyst bed and the distribution tray. Vendors provide loading guidelines and catalyst-handling contractors have developed innovative ways for loading this material.



1. Wait until the liquid feed reaches the bottom section of the first catalyst bed, as indicated by increasing temperatures in the bottom thermocouple array.
2. Manually open the first quench valve by at least 50%.
3. Wait until the liquid feed reaches the bottom section of the second catalyst bed, as indicated by increasing temperatures in the bottom thermocouple array.
4. Manually close the first quench valve and at the same time open the second quench valve by at least 50%.

This operating sequence should be repeated if there are more catalyst beds. If the exotherm becomes uncontrolled, cut the liquid feed.

FIG. 10. Controlling the pre-wetting exotherm.

Restart. Standard preparations for startup are commenced after the top elbow(s) are reinstalled on the reactor(s) and the reactor circuit is put into its startup configuration. All equipment opened or worked on should be handed back from engineering or maintenance following documented close-up and hand-over procedures, and a detailed (piping and instrumentation diagram) P&ID-based check of the condition of the unit is performed, including a cross-check that all blinds are in their correct position.

The reactor circuit is freed of oxygen—either by evacuation or nitrogen purging—prior to rolling the recycle gas compressor. When recycle gas circulation has been established, the heater is commissioned and the system pressure is raised to 25% of normal operating pressure, with hydrogen (H_2), until the minimum pressurization temperature (MPT) is reached.

During this heating/hold phase, when the mass of the reactor and catalyst is being brought to the MPT, the catalyst is being dried out. Any moisture the catalyst has adsorbed during transport, storage and loading is slowly driven off in this drying step. Normally, by the time the reactor has reached MPT, the catalyst dry-out has been completed; however, if this step is incomplete, the catalyst temperatures must be held to complete the dry-out.

Once all reactor skin temperatures are above the MPT, the system can be pressured to close to normal operating pressure and temperatures can be adjusted (as determined by the catalyst vendor) for catalyst sulfiding. During this process, pressure testing and leak detection steps are completed to ensure the circuit is tight before introducing liquid feed and/or hydrogen sulfide (H_2S).

In liquid-phase sulfiding, the startup oil—normally a low endpoint (< 370°C) diesel—is introduced at 105°C–110°C. During the introduction of the startup oil, called pre-wetting (**FIG. 10**), there will be an exotherm in each catalyst bed caused by the heat of adsorption as the liquid contacts and adsorbs onto the surface of the catalyst. Type-II hydrotreating catalysts must be kept below 140°C–150°C until the catalyst is fully wetted to retain the hydrotreating catalyst structure and activity. The temperature limit for hydrocracking catalyst (prior to the introduction of H_2S) is 205°C to prevent reduction of the metal oxides on the catalyst to their base metallic (zero valence) state in the H_2-rich conditions. Any reduced metal sites cannot be regenerated and result in permanent catalyst deactivation.

Traditionally, no quench has been required to maintain the catalyst temperature within the 140°C–150°C and 205°C limits mentioned above; however, as

the hydrocracker catalyst activity has increased in recent years (with the inclusion of zeolitic base material to hydrocracking catalysts), the heat of adsorption exotherms have increased. In addition, if the startup has been delayed and the catalyst subjected to an extended period (several days) of recycle gas-only circulation without any liquid feed, the wetting exotherm can be very high. In gas-phase sulfiding, the catalyst will be extremely dry by the end of the sulfiding step and pre-wetting exotherms can be extreme and must be proactively controlled with quench.

Therefore, the wetting exotherm can be significant, > 40°C for hydrotreating catalyst and > 100°C for hydrocracking catalyst, hence the low oil cut-in temperature. An additional factor that must be understood is that the wetting exotherm is extremely fast in its increase and subsequent decrease—much faster than the onset of a reactor catalyst temperature runaway during normal operation. If the reactor(s) has temperature rate-of-change alarm/trip functionality, a high probability exists of it being activated during catalyst pre-wetting. This very severe exotherm must be controlled proactively. If the operator waits until the exotherm starts in a particular catalyst bed, it is too late to control it with quench. Therefore, it is imperative to start the quench flow a few minutes before the liquid reaches the top of the respective reactor/bed. For example, in a multi-bed reactor, as the exotherm reaches the first thermocouples in the bottom of the first bed, the quench between the first and second beds should be opened (to at least 50% of the quench valve capacity) and the quench to the current bed decreased and closed. This process should be repeated for each bed (**FIG. 10**).

After pre-wetting and catalyst flushing are complete, the catalyst temperatures can be increased for low-temperature catalyst sulfiding (normally 230°C). At this time, the H_2 purity in the recycle gas should be above 90 vol%, ideally, and there should be no wash water or amine circulation.

Catalyst conditioning. This consists of sulfiding of the active metal sites on hydrotreating and hydrocracking catalysts and is the most important and most sensitive step in the startup procedure. Most hydrotreating and hydrocracking catalysts have metals—nickel, molybdenum, cobalt and tungsten are most prevalent—that are in the oxide state when manufactured and must be converted to the sulfide state to make them effective and stable for hydrotreating and hydrocracking reactions. The sulfiding reactions include the following (Eqs. 1–3):

$$3NiO + H_2 + 2H_2S \rightarrow Ni_3S_2 + 3H_2O + heat \quad (1)$$

$$MoO_3 + H_2 + 2H_2S \rightarrow MoS_2 + 3H_2O + heat \quad (2)$$

$$WO_3 + H_2 + 2H_2S \rightarrow WS_2 + 3H_2O + heat \quad (3)$$

If the catalyst has been presulfided or pre-sulfurized, the startup process is much simpler, using a procedure provided by the vendor. For presulfided catalysts, the above reactions have been completed using a carefully controlled ex-situ procedure following manufacture; therefore, the active metals are already in their sulfided form. Startup comprises heating in a controlled manner in a H_2 environment.

Presulfurized catalyst have a sulfur-rich compound (normally a polysulfide) bound to the catalyst that decomposes at a known temperature and provides sulfur at the active metal sites during the startup. This eliminates the need to inject a sulfur compound, such as dimethyl disulfide (DMDS), into the reactor circuit. For presulfurized catalyst, the startup procedure is like that for catalyst, where the active metal sites are in the oxide form but simpler and so a little quicker. The startup of presulfurized catalyst is vulnerable where the startup is interrupted shortly after the decomposition of the sulfur-rich compound. If the reactor circuit must be depressured before the high-temperature soak, a loss of sulfur may require replacement by DMDS injected into the high-pressure circuit.

In-situ sulfiding can be done in a gas-phase procedure or in a procedure that uses liquid feed to the unit, normally on full recycle mode. To enable an efficient catalyst sulfiding step, a sulfur compound is injected into the high-pressure circuit, either directly or with the liquid feed, to provide the sulfur for the sulfiding reactions. The most common sulfiding compounds are dimethyl disulfide or DMDS (CH_3SSCH_3) and Sulfrzol 54 ($C_4H_9S_4C_4H_9$), shown in Eqs. 4 and 5:

$$\text{DMDS: } CH_3SSCH_3 + 3H_2 \rightarrow 2CH_4 + 2H_2S \quad (4)$$

$$\text{Sulfrzol 54: } C_4H_9S_4C_4H_9 + 11H_2 \rightarrow 8CH_4 + 4H_2S \quad (5)$$

DMDS is preferred due to its high-sulfur content, low decomposition temperature and low vapor pressure.

Gas-phase sulfiding has been replaced by liquid-phase sulfiding, especially since the development of Type-II hydrotreating catalysts. As mentioned above, the high-activity Type-II hydrotreating catalysts have a temperature limit of 140°C–150°C prior to being fully wetted. To allow gas-phase sulfiding, a protective compound must be applied to the catalyst before delivery and loading to allow the catalyst to be fully sulfided at temperatures up to 315°C–320°C before liquid feed is cut in.

At 200°C, the injection of the sulfiding agent can be started. Increasing the injection rate should be done carefully and in a step-wise manner to avoid excessive exotherms from the decomposition of the sulfiding agent and the initial sulfur laydown. The injection rate can be kept at the maximum (set by the catalyst vender) until H_2S breakthrough is achieved, unless excessive catalyst exotherms are experienced, at which time the injection rate must be reduced until the exotherms return to within acceptable limits.

Regular sampling and composition testing of the recycle gas, low-pressure separator and fractionator off-gasses, and liquid streams should commence with the start of catalyst sulfiding to monitor the progress and allow the calculation of a sulfur balance. Shortly after sulfiding is begun, water will begin to accumulate in the cold high-pressure separator and will continue until sulfiding is complete.

Low-temperature sulfiding conditions should be maintained until H_2S breakthrough is achieved and at least 70% of the theoretical sulfur up-take has been injected in the form of the sulfiding agent. When these criteria have been met, the catalyst temperatures can be increased for high-temperature sulfiding.

Through the catalyst sulfiding process, recycle gas H_2 purity decreases as H_2 is consumed by the decomposition of the sulfiding agent and the only H_2 make-up is to compensate for solution losses. If the H_2 purity drops to 70%–75%, the high-pressure bleed should be opened to purge some recycle gas and allow a higher H_2 make-up. This will result in H_2S loss from the recycle gas and should be minimized.

As the H_2 purity drops and the methane (CH_4) content increases through the sulfiding process, the density of the recycle

gas increases significantly and the load on the recycle gas compressor driver also increases. For an electric motor-driven recycle gas compressor, the load on the motor (amps) should be monitored and the H_2 purity increased, if necessary—at the expense of H_2S loss—to ensure that the compressor does not trip on motor overload.

After H_2S breakthrough has occurred and before catalyst temperatures are increased, the injection rate of the sulfiding agent should be decreased by about 30%–50% to keep the H_2S content in the recycle gas in the 1 vol%–1.5 vol% range. If the H_2S content goes too high, the H_2S losses resulting from solution losses become excessive. As the catalyst temperatures increase, the rate of sulfur laydown on the catalyst increases again and the injection rate of the sulfiding agent can be increased. The high-temperature sulfiding soak for hydrotreating catalyst is normally 315°C–320°C and 290°C for the hydrocracking catalyst.

When increasing catalyst temperatures from the low-temperature sulfiding step in the transition to high-temperature sulfiding, an additional precaution is required for highly active hydrocracking catalysts (higher zeolyte content). An injection of a nitrogen (N)-based modifier stream (ammonia or high-N liquid feed) is required before the catalyst reaches 260° to avoid the impact of excessive cracking of the circulating liquid and coking of the fresh, hyperactive hydrocracking catalyst. Production of significant quantities of liquefied petroleum gas (LPG) is a sign that hydrocracking reactions are occurring: the temperature increase should be stopped and the injection rate of the nitrogen-based inhibitor increased.

The high-temperature soak period (normally 4 hr–8 hr) commences after the catalyst temperatures reach the designated high-temperature sulfiding temperatures. The H_2S content in the recycle gas should be controlled in the 1 vol%–2 vol% range. When this period is complete and when catalyst bed temperatures have stabilized (no exotherm), no more water is being produced and the injection rate of the sulfiding agent has reduced to the level of the H_2S losses via solution losses (recycle gas H_2S content stable or increasing slowly), the catalyst sulfiding step is deemed to be complete. The sulfur balance should also confirm the completion of sulfiding.

A catalyst sulfiding process that is smooth and relatively fast normally creates the best results. Time wasted between the various startup steps results in a higher probability of unwanted outcomes, such as running out of sulfiding agent or having the catalyst at elevated temperatures (> 200°C) for too long (hr) before injecting the sulfiding agent.

After completing gas-phase sulfiding, the catalyst temperatures are reduced for the cut in of liquid feed (see the discussion above on catalyst pre-wetting). If liquid feed has already been established, normally on circulation, the catalyst temperatures should be reduced for the switch to normal feed quality: straight run vacuum gasoil (VGO) or diesel for a hydrotreating unit, or VGO for a hydrocracker. Normally, cracked or high endpoint feedstocks are not permitted for 3 d–7 d to allow the catalyst to stabilize without excessive coke laydown on the fresh catalyst.

After completing liquid-phase sulfid-

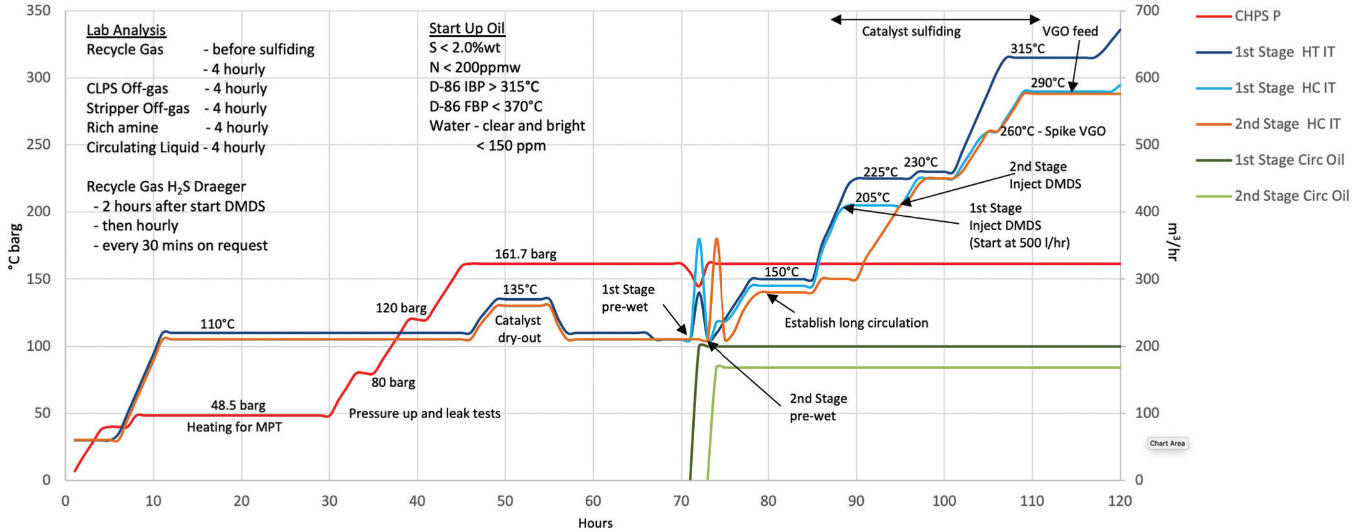


FIG. 11. Two-stage hydrocracker startup plots.

ing in a hydrocracking unit, the circulating startup oil (diesel) will be completely desulfurized and switching to VGO will result in a significant and rapid temperature rise in the hydrotreating catalyst beds. VGO can be cut in over a period of 3 hr–5 hr at constant catalyst bed inlet temperatures (determined by the catalyst vendor) and in incremental tranches to control the transition.

During this transition there should be no hydrocracking conversion activity. The hydrocracking catalyst will be sensitive to the high organic N-slip from the hydrotreating catalyst until target hydrotreating temperatures and hydrodenitrogenation (HDN) activity are attained. The organic N-slip will temporarily passivate the hydrocracking catalyst, reducing its initial activity. The aim, once the liquid feed has been switched to 100% VGO, is to increase the hydrotreating catalyst temperatures quickly, within the constraints set by the catalyst vendor, to establish the design N-slip to the hydrocracking catalyst while keeping the hydrocracking catalyst temperatures as low as possible.

When design conditions have been reached at the outlet of the hydrotreating catalyst, the hydrocracking catalyst temperatures can be increased carefully. A descending temperature profile should be adopted in the hydrocracking catalyst beds (each catalyst bed inlet temperature lower than the previous bed inlet). Once conversion has been established, the feed rate can be increased to 70%–75% of design and the fractionation section lined out. Over time, during this period, the organic N will be stripped from the hydrocracking catalyst and its activity will slowly increase. This additional activity will allow a higher conversion, or a higher feed rate at constant conversion.

Two-stage hydrocracking unit. For a two-stage hydrocracking unit, two approaches to the operation of the second stage are available once VGO is introduced. The second stage of a two-stage unit normally operates in an almost N-free environment except for a small organic N-slip from the first stage (< 5 ppmw) and a quantity of ammonia in the recycle gas (higher if the unit does not have a recycle gas scrubber).

As mentioned above, the organic N-slip from the first stage will be high until the hydrotreating catalyst reaches its target temperature and HDN activity. If the hydrocracking catalyst can tolerate the high initial organic N-slip, the approach outlined above can be followed. However, if it is deemed inappropriate to subject the second-stage catalyst to high levels of organic N, then the feed can be removed and the catalyst temperatures reduced to < 260°C, until such time as the organic N-slip from the first stage is within specification (normally 12 hr–24 hr). When the second-stage feed (recycle oil) is nitrogen-free, it can be reintroduced and the second stage conversion established, as above.

FIG. 11 provides an example of the predicted behavior of the main process variables of a two-stage hydrocracker for a typical startup procedure (liquid-phase sulfiding), as described above.

The catalyst vendor will suggest temperature limits on the operation of the hydrotreating catalyst as well as the conversion levels for the first few weeks of operation. During this initial period of operation, it is beneficial to limit the severity of operation (no difficult refractory feeds or full conversion) until the catalyst has stabilized. These short-term limitations, for a few weeks, can add several months to the catalyst cycle. HP

NOTES

[a] Petroval's S-Load system

ACKNOWLEDGEMENT

The authors want to thank Trevor Penny from CR International for the images and helpful comments.

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LEE NICHOLS, Editor-in-Chief/Associate Publisher

The 1990s: Clean fuels and emissions mitigation, M&A, GTL and the fieldbus wars

Much like several initiatives passed in the 1970s and 1980s, the 1990s were a decade heavily focused on environmental issues, with many new regulations being enacted to not only mitigate industrial and vehicle emissions but also to advance the production of clean fuels globally. As a result, refiners spent billions of dollars during the 1990s to install, modify, upgrade and reconfigure process units to adhere to new government regulations [e.g., the establishment of the reformulated gasoline program (Phases 1 and 2) in the U.S.].[222] This trend is still progressing today. Additional greenhouse gas emissions reduction initiatives also emerged from the Kyoto Protocol in the late 1990s/early 2000s, a precursor to the Paris Agreement in 2016. Both agreements call upon nations to significantly mitigate carbon emissions.

New clean-fuels regulations led to additional refining capacity being built during the 1990s to reduce sulfur levels in transportation fuels. In conjunction with new secondary unit capacity builds, the refining industry increased total net crude distillation capacity by nearly 8 MMbpd, reaching more than 83 MMbpd by 2000.[181] Although rocked by an economic crisis in the late 1990s, the Asia-Pacific region led refining capacity additions, adding more than 8 MMbpd (net) by 2000—China alone more than doubled domestic refining capacity to nearly 6 MMbpd within the decade.[181]

The 1990s also witnessed the increased usage of metallocene catalysts. Metallocene structures were first discovered simultaneously in the 1950s by Thomas Kealy and Peter Pauson at Duquesne University (U.S.), and by a different group comprised of Samuel Miller, John Tebboth and John Tremaine at British Oxygen (now part of Linde) in London, England—these groups worked with ferrocene, a type of metallocene.[223, 224]

A commercial use for metallocene was discovered in the 1970s by German chemist Walter Kaminsky while working at the University of Hamburg (Germany).[225] According to literature, Kaminsky discovered that using metallocene with a methyl aluminoxane cocatalyst led to a novel pathway for olefin polymerization. This discovery led to many companies increasing their research and development budgets to produce new metallocene catalysts for polymer production (e.g., polyethylenes, polypropylenes, polystyrene). Many companies introduced proprietary metallocene catalysts in the 1990s, including ExxonMobil, Dow Chemical, BASF and Mitsui, among others.[226]

The 1990s also witnessed a surge in high-profile mergers and acquisitions, the advancement of gas-to-liquids (GTL) technologies and capital-intensive GTL plant builds, and the evolution of fieldbus technologies and the standards that govern them.

New fuel standards directives lead to a decline in sulfur content. One of the detrimental effects of modernizing societies is an increase in air pollution. Increased smog from fuel exhaust has been a challenge in many cities around the world for more than 70 yr. Many nations' governments have enacted a host of regulations and standards to combat air pollution. For example, the U.S. began to enact new air pollution laws in the mid-1950s. These directives led to the Clean Air Act and various amendments, which gave more authoritative power to the U.S. Environmental Protection Agency (EPA) to mitigate air pollution. These acts/amendments also created federal standards for vehicle emissions (the history of the Clean Air Act and subsequent amendments are detailed in the June issue's History of the HPI section).

The Clean Air Act of 1990 led to the creation of tiered emissions standards for vehicles in the U.S. The Tier 1 standard was introduced in 1991 and phased into the market between 1994 and 1997. Tier 2 was adopted in 1999 and phased-in from 2004–2009.[227] The Tier 2 program marked the first time the U.S. EPA treated vehicles and fuels as a system.[228] These standards applied to light-duty vehicles (i.e., vehicles ≤ 8,500 lb) and led to a significant decline in sulfur limits in transportation fuels (gasoline and diesel). Prior to these regulations, on-road diesel fuel sulfur content was more than 5,000 parts per million (ppm). To adhere to new emissions standards put forth in the Tier 1 program, new low-sulfur diesel fuel was introduced into the market in the early 1990s. This fuel met the new sulfur limit specification of 500 ppm, significantly decreasing total sulfur content in diesel fuel.[229]

The Tier 2 standard helped reduce sulfur content in gasoline by up to 90%. By early 2004, the corporate average gasoline sulfur standard was 120 ppm, with a cap of 300 ppm. This standard was reduced to 30 ppm with an 80-ppm cap in 2006.[230] The Tier 2 standard also reduced sulfur content in diesel fuel to 15 ppm, which became known as ultra-low-sulfur diesel (ULSD). Subsequent regulations would further decrease sulfur content in fuels—the Tier 3 fuel regulation, adopted in 2017 with implementation to 2025, further reduced sulfur content in gasoline to 10 ppm.

In Europe, research by French and German scientists into smog mitigation in major cities in the mid-1950s led to the origins of new emissions standards

in the region. Their work led to Directive 70/220/EEC in 1970, which was the impetus to setting emissions standards for light- and heavy-duty vehicles in Europe (this research was detailed in the History of the HPI section of the June issue).[145] This directive eventually led to the introduction of the Euro 1 standard in 1992 (implemented for passenger cars in 1993), the removal of leaded petrol from filling stations in Europe and the adoption of three-way catalytic converters.[144,146,147] The Euro 1 standard was replaced by Euro 2 in 1996, which reduced sulfur limits in diesel from 2,000 ppm to 500 ppm; Euro 3, introduced in 1999 and implemented in 2000, further reduced sulfur content in diesel to 350 ppm and gasoline to 150 ppm. Subsequent standards (i.e., Euro 4–6 and Euro I–IV) would continuously reduce sulfur content in transportation fuels to near-zero levels.

The implementation of European emissions standards in the early 1990s would eventually become a global standard for many countries around the world to adhere to new clean fuels regulations. For example, more than a dozen Asian nations started using European emissions and fuel quality standards in the late 1990s/early 2000s. Many still use these standards as a benchmark for sulfur content in transportation fuels. The same initiatives apply in other regions, such as Africa, Central and South America, and the Middle East.[231]

These new sulfur cap limits in transportation fuels have had significant impacts on refiners globally. To produce fuels that adhere to these standards (e.g., Tier 3 in the U.S., Euro 6/VI in Europe and other parts of the world), refiners must invest a significant amount of capital in new secondary units. Since the adoption of emissions and sulfur content standards in the U.S., Europe and other nations, tens of millions of barrels per day in new secondary unit capacity have been built at a cost of hundreds of billions of dollars.

Consolidation in the oil industry: Mergers and acquisitions that created mega-companies. The 1990s witnessed several significant mergers and acquisitions that created some of the largest integrated companies in the world. Combined, these deals exceeded $220 B and included the following:

- **bp and Amoco:** In 1998, bp merged with Amoco in a more than $48-B deal, which was the largest industrial merger ever up to that point in time.[232] The merging of the two companies created an energy conglomerate with a market capitalization of $110 B. The deal was complementary for both sides. bp, through Amoco, strengthened its refining and chemicals production and products marketing—Lord Brown (bp's chief executive 1995–2007) said that through the merger, bp gained 9,300 fuel stations and five refineries that produced a total of 1 MMbpd.[233] By merging with bp, Amoco gained a foothold on the international market, a weak spot for the company at the time.
- **bp Amoco and ARCO:** Not even 1 yr after bp and Amoco merged, bp Amoco acquired the Atlantic Richfield Co. (ARCO) for $27 B. The acquisition significantly increased bp Amoco's foothold in Alaska's North Slope (U.S.) oil exploration and production operations, as well as captured 20% of California's (U.S.) fuels retail market—at the time, ARCO owned approximately 1,200 fuel service stations in the state.[234]
- **Total and Petrofina:** In 1998, Total acquired Belgian oil company Petrofina for $12 B. The deal created the third-largest company (TotalFina) in Europe, and the sixth-largest company in the world. The acquisition helped Total gain a more international foothold, increased the company's refining and products marketing operations, and provided the new organization (TotalFina) with a market capitalization of nearly $40 B.[235]
- **TotalFina and Elf Aquitaine:** Approximately 8 mos after acquiring Petrofina, TotalFina acquired Elf Aquitaine (Elf) for approximately $54 B.[236] At the time, Elf was a major integrated oil and gas company and one of the largest petrochemical companies in the world. TotalFina not only gained sizable exploration and production operations in West Africa and the North Sea from Elf, but also its petrochemicals and chemicals production capacity, five refineries and Elf's 6,500 fuel stations throughout Europe and West Africa.[237] After the merger was completed, TotalFina Elf became the world's fourth-largest company.[237]
- **Exxon and Mobil:** In 1998, Exxon announced an $81-B deal to merge with Mobil, which would create the third-largest company in the world behind General Electric and Microsoft.[238] The U.S. Federal Trade Commission (FTC) unanimously approved the merger in late 1999 dependent on the two organizations agreement to divest a sizable amount of assets. For example, the FTC ordered the two companies to sell more than 2,400 fueling stations in the northeast U.S., California and Texas; Exxon had to sell its refinery in Benicia, California, and agreed to stop selling gasoline and diesel fuel under the Exxon name in the state for 12 yr; and other assets. These demands from the FTC were the largest divesture ever asked by the commission up to that time.[239] The merger not only created a mega-company with a market capitalization value of nearly $240 B, but also reassembled two pieces of John D. Rockefeller's Standard Oil empire that was broken up in 1911.
- **Shell and Texaco:** In 1997–1998, Shell and Texaco agreed to a partial merger of downstream operations and fuel stations in the west and Midwest portions of the U.S. The JV, Equilon Enterprises, operated eight refineries, 10 lubricant plants, more than 70 oil and product terminals, and more than 11,200 fueling stations and convenience stores.[240] Equilon soon joined Saudi Refining (now Saudi Aramco) to create Motiva Enterprises, which eventually would operate one of the largest refineries in the world, the 630,000-bpd Port Arthur refinery in Port Arthur, Texas (U.S.). Shell would eventually retain all Equilon Enterprises and Texaco's share in Motiva to pave the way for Chevron and Texaco's $39-B merger in 2001.

Bintulu and Mossel Bay: The world's first GTL complexes. The first wide-scale use of synthetic fuels production from syngas was in Germany in the 1930s and early 1940s. These facilities utilized the Fischer-Tropsch (FT) process, a chemical reaction that converts carbon monoxide (CO) and hydrogen into liquid hydrocarbons (e.g., transportation fuels). According to literature, by the mid-1940s, Germany had nine plants in operation that used the FT process. Combined, these plants produced approximately 600,000 tpy of synthetic fuels.[241]

FT synthesis was the basis for several plants developed by Sasol in South Africa beginning in the mid-1950s. These included Sasol-1–3 plants, which used coal as the primary feedstock, later transitioning to natural gas in the early 2000s—Sasol developed and commercialized its Slurry Phase Distillate FT process at the Sasol-1 plant in Sasolburg, South Africa in the early 1990s.[242] Sasol-2 and Sasol-3 plants were built in the early 1980s as a direct effect of the oil crises of the 1970s—these plants accounted for $6 B in capital investments and utilized proprietary GTL technology from Sasol (the global oil crises of the 1970s were detailed in the History of the HPI section of the June issue).[241]

Sasol's FT technology was then utilized for the Mossgas GTL plant, which, upon completion in 1992, became the world's first commercial-scale GTL plant using natural gas as a raw material for syngas production.[241] The Mossgas GTL plant, located in Mossel Bay, eventually fell into the hands of The Petroleum, Oil and Gas Corp. of South Africa (PetroSA)—the national oil company of South Africa—after its formation in 2002 upon the merger of Soekor, Mossgas and parts of the Strategic Fuel Fund Association.[243] The facility converts natural, methane-rich gas into high-value synthetic fuels. According to PetroSA, the technology uses a series of conversions starting with the reforming of methane to carbon dioxide (CO_2), CO, hydrogen and water. The CO-to-hydrogen ratio is adjusted using the water-gas shift reaction and the removal of excess CO_2 in an aqueous solution of alkanolamine. The synthesis gas is then chemically reacted over an iron or cobalt catalyst to produce liquid hydrocarbons (gasoline, kerosene, diesel) and other byproducts.[244]

The Mossel Bay GTL plant has been expanded over the past two decades, reaching a total installed capacity of 36,000 bpd—a crude oil equivalent of 45,000 bpd. Sasol has also improved its FT-GTL process, which was used for Sasol's second large-scale GTL plant, Oryx GTL. The Oryx GTL facility—a JV between Sasol and Qatar Petroleum—in Ras Laffan City, Qatar started development in 2003 and began operations in 2007. The 34,000-bpd plant was built at a total cost of nearly $1 B. Sasol would later provide its FT technology to Chevron for the nearly $10-B, 33,000-bpd Escravos GTL plant in Escravos, Nigeria, and the $3.4-B, 1.5-MMtpy Oltin Yo'l GTL plant in Uzbekistan.

Shell was another company that devoted significant resources to the development of a GTL technology and subsequent capital-intensive investments in

FIG. 1. View of the Pearl GTL plant. Photo courtesy of Shell.

new GTL processing capacity. In 1993, Shell commissioned its first GTL plant in Bintulu, Malaysia; however, research on this processing technology took decades to complete. Shell started conducting research on FT processes in 1973 at its labs in Amsterdam, Netherlands. The company first focused on coal-to-liquids conversion but later switched to natural gas as the primary feedstock.[245] Within these tests, the company was able to create new catalysts to produce a few grams/d of hydrocarbon liquids from natural gas. By 1983, production increased to a few bpd at Shell's pilot plant facility in Amsterdam.[246]

Less than a decade later, Shell opened the Bintulu GTL facility. The $850-MM, 12,500-bpd plant utilized Shell's Middle Distillate Synthesis (MDS) process. According to literature,[247] the MDS process is comprised of three basic stages. These stages include:

- **Stage 1:** The production of syngas from the partial oxidation process of natural gas with pure oxygen via Shell's Gasification Process.
- **Stage 2:** The syngas passes through paraffin synthesis reactors equipped with proprietary Shell catalyst. These catalyst and reactors favor the formation of long-chained liquid molecules (wax), simultaneously minimizing the formation of gaseous compounds.
- **Stage 3:** The intermediate and waxy synthetic crude oil molecules are converted and fractionated into high-quality products. The waxes are purified via a hydrogenation unit followed by advanced fractionation. Clean middle distillates and waxy raffinate are produced by a selective hydrocracking process (i.e., heavy paraffin conversion), followed by distillation.

The Bintulu GTL plant was later expanded to 14,700 bpd in the mid-2000s. Several years later, Shell and Qatar Petroleum commissioned the largest commercial GTL plant in the world, the $18-B, 140,000-bpd Pearl GTL complex in Ras Laffan Industrial City, Qatar (**FIG. 1**).

The fieldbus wars lead to a new standard in process automation. In the mid-1970s, the introduction of the distributed control system (DCS) by Honeywell and Yokogawa revolutionized process automation in the refining and petrochemical industries (the history of the DCS is detailed in the History of the HPI section of the June issue). This advancement in automation led to several new technologies to optimize plant operations, including the creation of fieldbus.

According to literature, fieldbus is the technology that provides a digital link between intelligent, microprocessor-based field instrumentation and the host DCS.[248] Prior to fieldbus, field instruments had to be wired in a point-to-point configuration; fieldbus enabled these instruments to communicate with the DCS using a single wire.

The origins of fieldbus technology date to the mid-1970s with the creation of the general-purpose interface bus, a precursor to Intel Corp.'s Bitbus in the early 1980s. Throughout the 1980s, several companies developed fieldbus technologies for use in several different industrial applications, including process automation for the refining and petrochemical sectors. The proliferation of fieldbus technologies in the late 1980s–mid-1990s led to many different systems that were not compatible with competing technologies, and several international standards organizations fought for their fieldbus standard to be accepted by industry. This predicament—known as the fieldbus wars—led to many users trying to seek a unified standard that enabled them to utilize different/competing technologies (i.e., a plug-and-play solution).[248] The fieldbus wars included competing standards in Europe (e.g., the French FIP vs. the German PROFIBUS, which the two later tried to combine) in the late 1980s/early 1990s and within the U.S. in the mid-1990s.[249]

In 1999, the leading fieldbus manufacturers at the time—ControlNet, Fieldbus Foundation [developed by the International Society of Automation (ISA) and purchased by the FieldComm Group in 2015], Fisher Rosemount (now part of Emerson), the PROFIBUS user organization, Rockwell Automation and Siemens—signed an agreement that put an end to the fieldbus wars.[249] This agreement became the basis for the International Electrotechnical Commission's (IEC's) IEC 61158 standard. According to literature, the IEC 61158 standard grouped the different fieldbuses into types, but created common physical, data link and application layers.[248] The standard enabled competing technologies to work with each other. Fieldbus is still in use today; however, it is being challenged by industrial Ethernet, a technology that has gained prominence since the 2010s. HP

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Excerpts from the 1990s: Process optimization, fuel quality and environmental compliance

EnCs in year 2000 and beyond

R. L. Tucker, January 1990

The future is best described in three phases: Drivers for change, current responses and projections. This article details the drivers for change within the next decade (e.g., changing workforce, competition, shrinking world, emphasis on quality, nature of projects, changing client approaches and technology availability), how the hydrocarbon processing industry's engineering and construction companies are responding to these changes, and projections for the future.

Implementing advanced controls with new DCSs

C. R. Aronson and D. C. White, June 1990

In the past few years, a new generation of distributed control systems (DCSs) has appeared. The incentives for utilizing these systems are the increases in computational capabilities, the facility for providing single operator windows and the increased potential for plant process information network development. These article will address the following three questions regarding implementation of advanced control projects:

1. What is the proper distribution of control and calculation functions between the DCS and a process computer?
2. How should the operator interface for the advanced controls be implemented?
3. What are the important factors to consider for project implementation and long-term maintenance of such systems?

EC seeks gasoline emissions control

September 1990

Gasoline emissions control is being given priority within the European Community as part of an overall strategy to reduce pollution by photochemical ozone in the lower atmosphere.

Estimating hydrogen of FCC coke

R. Sadeghbeigi, February 1991

Historically, the hydrogen content of coke has been used to judge the performance of fluid catalytic cracker (FCC) reactor strippers. The hydrogen content in coke is determined by performing an oxygen balance around the regenerator. The purpose of this article is to:

- Discuss the significance of hydrogen in coke
- Illustrate steps to calculate the hydrogen content of total coke
- Provide a graphical solution for the determination of hydrogen in coke.

Analyzing process energy efficiency

J. H. Siegell, March 1991

There are numerous methods for evaluating process energy efficiency. They involve representations of the process system to varying degrees of complexity in the search for lost Btus. A method exists which can be used to easily establish energy targets and compare current operation to these targets. This article details a new method that establishes energy targets for comparison with current operation and identifies specific areas for improvement.

Developing countries face big energy needs

HPImpact, April 1991

Developing countries must have energy to raise productivity and improve their living standards, but supplying that energy raises serious financial, institutional and environmental problems, according to a report by the Office of Technology Assessment. The report states that the magnitude of these problems underlines the need for more efficient production, transformation, delivery and use of energy.

Maximize steam generation with gas turbine HRSGs

V. Ganapathy, October 1991

Combined with heat recovery steam generators (HRSGs), gas turbines can operate in combined-cycle or cogeneration modes, thereby improving the efficiency of the overall system compared to the operation of the gas turbine alone. The type of HRSG suitable for a given firing temperature, the amount of oxygen consumed or depleted for a given fuel input, and the efficiency of the HRSG system as a function of fuel consumed are discussed.

Improve corrosion control by computer simulation

P. R. Petersen, January 1992

In many processes in refineries and chemical plants, corrosion occurs in the areas where acidic gases and water vapor condense together to form dilute solutions of acids. Results of an extensive study into the chemistry and properties of the various common neutralizing amines used to combat acid corrosion, the acids that are neutralized and the salts that are formed are reported in the work. The first section will describe theoretical considerations, lab studies and the models that were developed from these studies. The second section will discuss some of the practical applications of this knowledge.

Develop a strategic automation plan

J. B. Stout, May 1992

The development of a strategic vision, and the resulting masterplan, are the key to a successful automation program. The masterplan serves as a compass, maintaining focus on the automation journey while enabling the company to respond tactically to emerging business issues. This article examines organizations, architectures, applications and benchmarking as components of a masterplan, which is key to a successful automation program.

Trends in hydrogen plant design

T. Johansen, K. S. Raghuraman and L. A. Hackett, August 1992

Steam reforming will continue to be the main source for hydrogen production. Understanding important design considerations for hydrogen production via steam reforming require detailed attention to the many elements that comprise the process. Design trends focus on improvements to the plant's three principal unit operations:

- Generation of hydrogen/carbon monoxide syngas
- Conversion of carbon monoxide in the syngas
- Separation/purification of hydrogen from syngas.

Improve activated carbon bed adsorber operations

G. C. Shah, November 1992

Activated carbon beds (ACBs) provide the final cleanup phase of most vapor streams before venting to air. Generally, ACB applications include volatile organic chemicals (VOCs) control, odor removal and recovery/recycle of hydrocarbons.

Properly designed ACBs have a long service life; however, they do develop problems. This work provides a few troubleshooting tips to solve ACB poor performance situations.

Safety system performance terms: Clearing up the confusion

P. Gruhn, February 1993

The problem of ambiguous terms is particularly acute when dealing with safety systems. These systems are often referred to as emergency shutdown systems or interlock systems. For example, ask a chemical engineer and a safety engineer the meaning of "failure," "availability," "reliability," or "integrity" and you will probably get surprisingly different answers.

The problem is real and is causing considerable confusion in the safety control industry. This work will discuss how the industry can improve reliability and performance by using the right terminology.

Process optimization in the HPI

O. Pelham, July 1993

The increasing complexity of refineries and petrochemical plants requires closer integration and cooperation between traditional licensing activities and newer process control and optimization technologies. Greater complexity is being driven by new business, environmental and safety pressures. Advanced process control and information systems must accommodate expected changes in process configurations.

Single-sited catalysis leads next polyolefin generation

A. A. Montagna and J. C. Floyd, March 1994

New catalyst families are broadening product opportunities for polyolefins. An updated catalyst system, metallocene, has entered the technology race. This catalyst offers single-site activity, which narrows molecular weight and gives molecular weight distribution control for ethylene and propylene-based polymers. A comparative history details how metallocene-based polymers stack up to conventional Zieglar-Natta catalyst-produced resins.

New strategies maximize paraxylene production

J. J. Jeanneret, C. D. Low and V. Zukauskas, June 1994

Strong consumption growth and the shutdown of some capacity in 1992 have eliminated the surplus of paraxylene (PX) capacity, which existed from 1990–1993. PX supplies are becoming tight and market prices have risen dramatically over the last several months, sparking considerable interest in new PX production capacity. However, adding new PX capacity does not mean building new grassroots units. Additional capacity can be "found" by creative use of existing benzene, toluene and xylenes resources. By making the most of existing facilities, producers can capitalize on the current upswing of the PX market cycle.

Build pollution prevention into system design

P. P. Radecki, D. W. Hertz and C. Vinton, August 1994

Imagine using a computer-based tool that allows process designers to find and include pollution-prevention information and technology during the conceptual design stage. Emerging computer-based tools are now incorporating environmental considerations into process development.

Simplify process hazard reviews with 3D models

G. Tolpa, October 1994

Replacing plastic models with 3D electronic versions of the facility helps hazard and operability review (HAZOP) teams to increase review efficiency, reduce meeting time and record/document critical changes made during procedural meetings.

Recycle plastics into feedstocks

H. Kastner and W. Kaminsky, May 1995

Thermal cracking of mixed-plastics wastes with a fluidized-bed reactor can be a viable and cost-effective means to meet mandatory recycling laws.

Use reliability-centered maintenance to identify real-world risks

R. B. Jones, October 1995

Using reliability-centered maintenance (RCM) methods, HPI companies can make realistic, tangible progress toward reducing risk, improving productivity, minimizing downtime and increasing profitability. Because the HPI uses complex processes and operational practices to manufacture products, an effective analytical technique such as RCM is needed to help identify and mitigate high-risk situations that may occur.

Use alloys to improve ethylene production

S. B. Parks and C. M. Schillmoller, March 1996

Selecting suitable cast heat-resisting alloys for ethylene furnace tubing can cost-effectively improve product yield and selectivity, reduce downtime and lengthen production runs. Using better alloys for tubing has enabled raising temperatures,

shortening residence time and lowering pressure drop in the cracking coils. The results are increased ethylene product selectivity and yields.

EnCs adopt matrix management

J. G. Munisteri, June 1996

After a decade of restructuring and reengineering, engineering and construction (EnC) companies have discovered that the traditional corporate management system is no longer viable. Many EnC companies have adopted a matrix management as the management system of choice in effectively controlling decentralized operations. This article explores what a matrix management system is, why it is used and what it can accomplish.

Control contaminants in olefin feedstocks and products

J. A. Reid and D. R. McPhaul, July 1996

To be competitive, olefin manufacturers must use low-cost feedstocks, which frequently contain contaminants. Equally important, olefin customers, who are using newer technologies, are specifying more stringent limits on contaminants when purchasing products. These contaminants affect products and catalyst systems, hinder operating processes and impair equipment for both the manufacturers and customers.

An overview of current process designs and technologies shows several cost-effective options to reduce or remove feedstock contaminants such as carbon monoxide, carbonyl sulfide, carbon dioxide, hydrogen fluoride, ammonia methanol and phosphine.

Estimate product quality with ANNs

A. Brambilla and F. Trivella, September 1996

Artificial neural networks (ANNs) have been applied to predict catalytic reformer octane number and gasoline splitter product qualities. Results show that ANNs are a valuable tool to derive fast and accurate product quality measurements and offer a low-cost alternative to online analyzers or rigorous mathematical models.

Six critical management issues

D. M. Woodruff, December 1996

From September 1995–September 1996, a survey of critical management and workplace issues was conducted. Executives and managers were asked to list the top three issues facing their organization. The top six critical workplace issues included:

1. People issues
2. Cost and competition
3. Government regulations
4. Leadership and management
5. Change and technology
6. Quality and productivity.

Detect corrosion under insulation

M. Twomey, January 1997

Real-time x-ray technology is available to detect corrosion quickly and reliably under the insulation of piping systems without disturbing the insulation. This technology has permitted many refineries and petrochemical facilities to successfully detect this problem in an expeditious and cost-effective manner.

Economically recover olefins from FCC offgases

D. Netzer, April 1997

The concept of ethylene and propylene recovery from FCC offgases is not new; however, its application has been infrequent. Here, two approaches to olefins recovery are addressed. In the first, ethylene is recovered as a dilute gas at a concentration of about 15 vol% and serves as raw material for ethylbenzene and, subsequently, styrene. In the second approach, ethylene is recovered as a pure polymer-grade liquid. Propylene recovery is identical for both approaches.

Using mass meters for liquid measurement

K. J. Haveman and M. C. McGhee, July 1997

Mass-base flow measurement is rapidly being accepted by users in a wide variety of industries. The ability of Coriolis mass meters to handle an extensive range of applications and fluids is one reason for the swift growth in this technology. This article provides the basic theory of Coriolis meters, why mass flowmeters are selected, the most successful applications, existing and potential uses in the petroleum industry, and proper meter installation.

What are Western Europe's petrochemical feedstock options?

S. Zehnder, February 1998

In Western Europe, petrochemical demand remains strong, in contrast to stagnant oil requirements. Under these conditions, refiners have new opportunities not only as a manufacturer but also as a petrochemical feedstock supplier. Tighter gasoline specifications are rejecting olefins and aromatics from local gasoline pools. Therefore, refiners can recover propylene and xylenes from processing streams for sale or feedstock purposes. These two petrochemicals have very strong consumption demands.

How much safety is enough?

R. P. Stickles and G. A. Melhem, October 1998

Current design codes and standards primarily seek to mitigate catastrophic consequences from process incidents—standards guide engineers on how to design equipment to avoid failures. Since the focus is on consequences, these design specifications may not fully address incident frequency and provide coverage for double or triple jeopardy. Consequently, companies that seek sound guidance to correctly assess cost and benefit from loss control measures must broaden the approach and consider risk-based evaluation techniques.

Outlook for global HPI surprisingly strong

HPImpact, November 1998

Despite the maturing of the petrochemical industry, world and regional outlooks for petrochemicals are strong. High five-year petrochemical growth rates are possible if the Asian financial crisis does not continue to worsen and push the beginning of the economic recovery well beyond 2000. The global industry is forecast to grow as much as 5%/yr to 2007.

Reduce maintenance costs with smart field devices

J. S. Masterson, January 1999

Intelligent microprocessor-based field devices rapidly replacing older types of instrumentation in refineries and petrochemical plants generate vast amounts of information that can be useful far beyond the control room. New smart field devices provide processing plants with a rich opportunity to improve processes and reduce costs.

To bid or not to bid

B. Lenehan, May 1999

Bidding for large international projects can be a huge gamble, with big stake money and no game rules. Presented here is a simple method of collating and comparing requests for proposals data in a ranking order for the tender selection committee (TSC) to respond to. The bid/no bid responsibilities lie firmly with the decision-makers in the TSC; however, by using uniform data, it can quickly make balanced and consistent decisions. Over time, the data collated can also be used for analyzing the accuracy of the decisions.

Refinery air quality enforcement issues

S. P. Hampton and D. D. Bradley, August 1999

Throughout the latter half of the 1990s, petroleum refineries have experienced an unprecedented level of enforcement actions relating to environmental compliance. Air quality issues are at the forefront of most refinery enforcement activity. By understanding the regulatory requirements and managing compliance, refineries can assess potential exposure before the next scheduled environmental inspection. By reviewing these issues, refineries can prepare for agency inquiries and address potential exposure areas.

Consider using hydrogen plants to cogenerate power needs

J. Terrible, G. Shahani, C. Gagliardi, W. Baade, R. Bredehoft and M. Ralston, December 1999

Many forces are reshaping the global hydrogen market; refiners have several options to receive steam and electrical power from a hydrogen plant. The supply of hydrogen, steam and electrical power by third-party specialists can be particularly valuable when these requirements are large enough to justify the development of an independent supply infrastructure to serve multiple customers. In several case histories, the varying scenarios that simultaneously produce hydrogen, steam and electrical power from a single production plant are discussed. The integration of a steam methane reformer with various power generation technologies such as a topping turbine gas turbine and condensing turbine are explored. HP

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Electric drive LNG can grow with the clean energy transition

As companies across markets set aggressive decarbonization goals and sustainability commitments to align with changing regulations and consumer demands for greener energy and fuel, the liquefied natural gas (LNG) industry is searching for ways to lower emissions.

Electric drive LNG (e-LNG) has emerged as a potential solution to help lower carbon emissions in liquefaction plants and, in some cases, cut costs. Familiar to the market and already commonly used in small- and medium-scale plants, e-LNG uses electric motors to drive refrigerant compression rather than combustion gas turbines. Of the 40 liquefaction plants the author's company has engineered, 26 are driven by electric motors. As early e-LNG adopters in the smaller scale segments have seen success and longevity, larger plants are beginning to look to electric drives for some or all operations.

E-LNG offers increasingly attractive—and in some areas, crucial—adaptability because its use of electricity allows it to be run by fossil fuels or renewables. This trait has led to its popularity in an oscillating energy market, where the costs and availability of renewable and fossil-based electricity are constantly changing.

However, as LNG asset owners, operators and investors increasingly eye this technology for their plants, they must understand it is not a one-size-fits-all solution. Significant development of clean energy and energy grid infrastructure will be required if e-LNG is to become the new industry standard.

Growing interest in sustainability. While many dynamics are at play in the proliferation of e-LNG, two factors stand out as clear drivers of the upward trend in adoption: decarbonization goals and economic benefits.

On the decarbonization side, the growing low-carbon and carbon-neutral LNG market is causing prices to fetch a premium. Most LNG achieves carbon-neutral status through carbon offsets, not through direct reductions in emissions. Liquefaction of natural gas contributes approximately 8%–10% of the total emissions along the LNG supply chain. As global interest in carbon-neutral LNG grows, and with end users looking to limit their Scope 2 and 3 emissions, the industry is searching for more direct decarbonization solutions as an alternative to relying on carbon offsets.

In late 2021, the International Group of Liquified Natural Gas Importers issued a framework of rules for declaring cargoes as carbon neutral. It outlines a series of steps, including the transparent sharing of greenhouse gas (GHG) emissions data and concerted efforts to reduce Scope 1, 2 and 3 emissions in LNG rather than relying on offsets, the primary method for pursuing carbon-neutral LNG production.

The decarbonization driver is strengthened by LNG companies' sustainability and decarbonization targets. Across industries, carbon footprint considerations are rising to the forefront of long-term planning; for industries already associated with fossil fuels, these commitments are an important aspect of maintaining a social license to operate.

The author's company published a report[1] that surveyed more than 490 leaders across 14 industries and found that 73% of respondents have set GHG reduction goals; this jumped to 89% among companies with more than $1 B in annual revenue.

The economic considerations go both ways. Separate from the emissions-reduction benefits of e-LNG, economic considerations have been integral in adopting the technology. However, these considerations go both ways—they have also been a deciding factor in cases when the technology was passed over in favor of gas turbines.

In some cases, e-LNG can lower liquefaction plant capital and operating costs (CAPEX/OPEX) because the motor is cheaper and requires less maintenance than gas turbines, resulting in more overall onstream time across the asset's lifecycle. This cost reduction tends to be a driver of electric motor adoption on projects with adequate and competitively priced available power. However, it would be inaccurate to say that e-LNG is cheaper overall.

Since e-LNG pulls significant power from the grid—especially in cases where a large production plant might be utilizing this technology—there are often cases in which the installation of electric motors requires significant upgrades to surrounding transmission infrastructure, the installation of a high-voltage substation, and more to ensure adequate and stable power. This is where any cost advantage of motor drivers will erode.

In such cases, it is often more cost-effective overall to choose gas turbines to drive LNG compression. As cost tends to be the deciding factor for any major infrastructure project, e-LNG is not a blanket solution. However, when combined with premium pricing for lower carbon LNG supply and lower cost renewable power, the incentive for e-LNG is expected to grow in the coming years to aid in decarbonizing the natural gas supply chain. Even if market conditions do not support e-LNG economics, adopting electric drives—considering fu-

ture market changes—may be advantageous to position a facility for a smoother transition to low-carbon operation.

The renewable barrier. If e-LNG compression is intended to lower Scope 1 and 2 emissions, the carbon intensity of the energy used to power the LNG compression train is essential. A plant that chooses an electric drive powered by fossil-based energy could claim it has no emissions at the plant itself, but its Scope 2 emissions would remain heavy.

For e-LNG to be truly carbon neutral, the motors must be powered by clean energy. In some cases, this becomes a barrier to adoption. Unfortunately, there are not enough renewables on the grid in many areas to power electric motors at the scale needed for a larger LNG production plant, and the available electricity is too expensive to make the investment worthwhile.

However, other areas have found success, casting a positive example for those plants with adequate energy resources. The west coast of Canada has many proposed projects pairing renewables with electric motors for LNG compression due to the significant availability of hydropower. As of 2019, Canada held more than 81,000 MW of hydroelectric power capacity, which provides a strong baseload of renewable energy to support these plants. Another potential route to decarbonization without grid renewables is carbon capture. Motors are still applied on the compressors with power supplied by an onsite power plant, but a centralized carbon capture unit can be installed instead of distributed across multiple point sources. Offtake for CO_2 product or infrastructure for sequestration is required near the terminal site to realize this as a feasible option.

Moving forward. Although the U.S., in many areas, lacks the renewable capacity to power large-scale e-LNG plants, it does have significant plans for the modernization and decarbonization of power. The Biden Administration's Infrastructure Investment and Jobs Act, passed in late 2021, provides billions of dollars for renewable development and upgrades to grid transmission infrastructure, enabling greater availability of low-carbon energy for future LNG plants.

As new LNG plants are established and older ones are expanded and upgraded, many look to electric drive as a sustainable solution. Though not a panacea, e-LNG offers an adaptable compression option that can grow with the clean energy transition, offering asset longevity as decarbonization targets are realized. HP

LITERATURE CITED

[1] Black & Veatch "Corporate sustainability goal setting and measurement report," 2021.

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LNG: Adapting to a more strategic role calls for effective quality analysis

Disruptions of global hydrocarbon supply chains have increased the importance of liquefied natural gas (LNG), causing larger numbers of buyers and sellers to consider it for satisfying energy demand. For some, this is unfamiliar territory, requiring greater knowledge of the technologies and practices involved, as billions of dollars will be changing hands.

One major aspect of LNG custody transfer involves basic questions of gas purity and energy content to ensure that the product can be added to local distribution networks without issues. This article will examine analyzer technologies to evaluate LNG through its production chain and to verify its suitability for pipeline use.

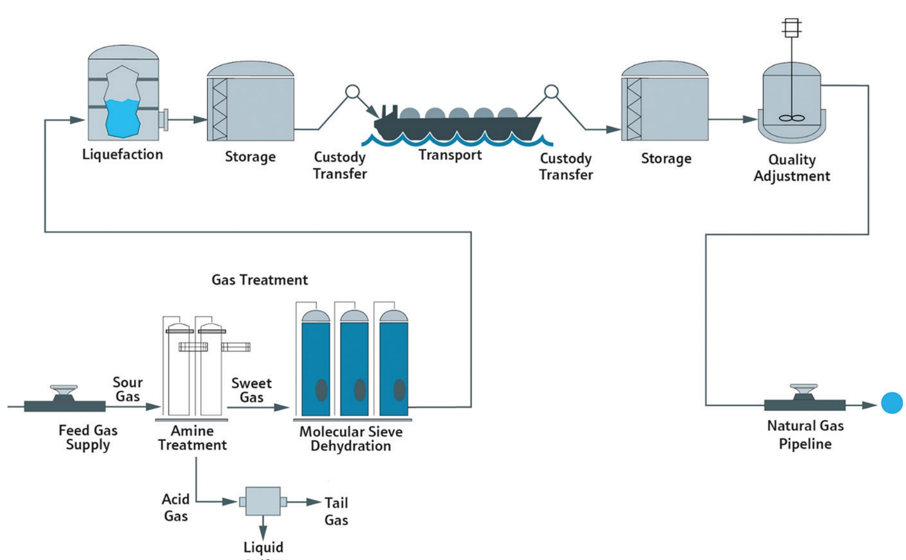
Under normal conditions, LNG is often sweeter and has fewer contaminants than locally produced sales gas available in most parts of the world. This is because feed gas goes through extensive amine and molecular sieve treatment to remove contaminants that can cause problems during the liquefaction process, such as hydrogen sulfide (H_2S), carbon dioxide (CO_2) and water (H_2O) (**FIG. 1**).

Additionally, heavier hydrocarbons must also be removed, since they condense at the temperatures involved. Nonetheless, it is possible to create off-spec LNG, as these contaminants can still be present, and, additionally, various diluents may remain, reducing calorific value and affecting the Wobbe index.

At this pretreatment stage, the contaminant levels involved normally call for tunable diode laser absorption spectroscopy (TDLAS) analyzers (**FIG. 2**) at transfer points between the stages to verify each step of the process. High-resolution TDLAS analyzers provide selective and specific on-line measurements of H_2S, CO_2 and H_2O in LNG feed gas.

Laser and detector components are isolated and protected from the process gas and entrained contaminants, thereby avoiding fouling and corrosion. This ensures stable long-term operation and accurate measurements in the field. Fast response to increases in H_2O concentration can trigger timely alerts when there is a breakthrough in molecular sieve adsorbent beds. Once pretreatment is complete, the gas moves to liquefaction and on to the transport mechanism.

In the transport stream. Once natural gas pretreatment is complete, it moves to the liquefaction stage, after which it



Description	Measured Component	Typical Range
Molecular Sieve Dryer Outlet	H_2O	0 – 10 ppm_v
Receiving Terminal	H_2O	0 – 10 ppm_v
Amine Treatment Unit Outlet	CO_2	0 – 100 ppm_v
Amine Treatment Unit Inlet	CO_2	0 – 500 ppm_v
Amine Treatment Unit Outlet	H_2S	0 – 10 ppm_v
Amine Treatment Unit Inlet	H_2S	0 – 500 ppm_v

FIG. 1. Natural gas must be treated before liquefaction to remove contaminants that can condense at cryogenic temperatures or damage equipment.

is transferred onto a ship for its primary journey. Once it arrives at a shoreside receiving terminal, the LNG is transferred to storage tanks for regasification before entering consumption pipelines. Some may be loaded to trucks and railcars while still in liquid form.

By the time any of this natural gas reaches an ultimate customer, it can change ownership multiple times, passing through custody transfer facilities at each handoff point. Various loads with potentially differing characteristics get mixed into the process, so there is no assurance that all loads are necessarily on-spec. LNG also changes while in transport, as it boils off its light components, with these components escaping into the tank or other enclosure. This alters the overall composition, and depending on the elapsed time and storage conditions, can significantly modify the calorific value of the load. Given the typical LNG tanker load can have a value of $50 MM or more, a change of even 1% is worth $500,000.

FIG. 2. TDLAS analyzers, such as the proprietary one shown[a], use differential spectroscopy techniques to quantify low parts per billion (ppb) to parts per million (ppm) levels of H_2O, H_2S and CO_2 in the outlet gas stream of an amine treatment unit.

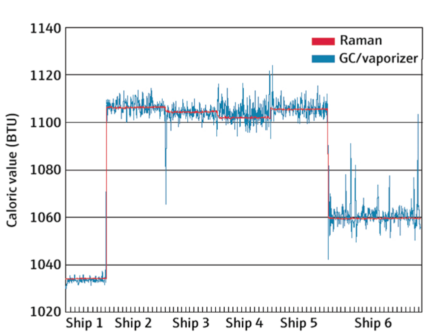


FIG. 3. Variability with the sample vaporizer results in erratic Btu readings from the GC.

Each time LNG passes through a custody transfer point, the new owner must verify its composition and calorific value characteristics to avoid paying for lost value or for passing off-spec product into the larger supply chain. This analysis does not need to be as stringent as during pretreatment processes, but basic contaminant levels and calorific values must be verified. The challenge is finding an analyzer technology suited to the application.

Gas chromatography (GC). The traditional method for analyzing natural gas after the pretreatment stage is with GC. Various models are available that can quantify the relevant components with the required precision and calculate the calorific value and Wobbe index. A typical GC analysis takes a sample of the natural gas, mixes it with an inert carrier gas and pushes it through a packed column enclosed in an oven. During the time in the column, the gases separate and exit individually, passing through a detector that measures the amount of each. Some models use multiple columns for a higher degree of separation.

Operationally, GC tends to be complex. First, these models depend on a sampling system to deliver gas to the analyzer. This requires tubing and valving over a limited distance, so the analyzer must be near the source. GC models often require a shelter, although some models can be installed in the open in forgiving environments. Second, GC models require a supply of consumables in the form of carrier gas and test gases for calibration, so there is an ongoing cost.

Third, while GC can perform this analysis routinely with sufficient precision, there is a weaker link in the analysis chain. An LNG sample must pass through a vaporizer to phase change the sample into a gaseous form suitable for GC analysis. The vaporizer stage is often more problematic than the analyzer itself, and in day-to-day operations, the vaporizer may not always provide a truly representative sample of the LNG—resulting in an inaccurate picture of the product's characteristics.

The task of the vaporizer is especially tricky because of the compositional complexity of natural gas. It must completely vaporize the sample without performing what is effectively a fractional distillation action. All the components do not vaporize simultaneously, so the vaporizer must ensure that it does not lose the lighter fractions or stop the process while some of the heavier fractions remain partially liquefied.

The vaporizer must operate within a very narrow range that is easily disrupted by changes in LNG flow, pressure or temperature. This makes it especially difficult to extract truly representative samples while the process is undergoing startup and shutdown actions. Vaporizer performance can be adversely impacted by changes in flowrate or changes in composition of the LNG. This can result in unstable vaporization, providing non-representative gas to the GC. Per the GIIGNL LNG custody transfer handbook, this data is not to be included in any CT calculations. Typical instability is visible as a noisy reading with random high and low spikes (**FIG. 3**), even when the supply being analyzed is very homogenous.

The overall sampling system—covering extraction, vaporization and transport to the GC analyzer—is highly complex, and all elements must work together to prevent pre-vaporization, incomplete vaporization or sample loss. This requires careful design, installation and maintenance throughout all steps, with precise temperature control. Given the procedure's difficulties, the motivation for simple and reliable alternative mechanisms is understandable.

Because natural gas is burned in gaseous form, it may seem counterintuitive to suggest evaluating its characteristics in liquid form, but its chemical composition and calorific value are not dependent on its phase. It may also seem counterintuitive to suggest using an analyzer at cryogenic temperatures. The solution is Raman spectroscopy, which can be applied to analyze natural gas whether in gas or liquid form.

Raman spectroscopy. Various gas and liquid analysis technologies have been developed around the ability of lasers to produce highly specific wavelengths of light. Various molecules are affected by this radiation at specific wavelengths, in characteristic and measurable ways, making it possible to detect and quantify chemical components of interest.

One method for determining vibrational modes of molecules is Raman spectroscopy, which uses a laser to produce light of visible or near-infrared wavelengths. When various molecules pass through this light, the light's energy excites the vibrations of molecular bonds, such as the bond between two hydrogen atoms. Inelastic scattering results in emissions of this excited energy, casting the laser light into different wavelengths.

What started as a single laser-generated color now becomes its own rainbow spectrum because each molecule in the sample produces a signal at a unique wavelength, and the relative intensity indicates the sample's molecular concentration. A Raman analyzer looks for these specific wavelengths and intensities to create a chemical profile of the sample.

When this concept is applied to LNG analysis, a probe is inserted into the pipe to analyze the flowing liquid or gas (**FIG. 4**). The probe is inserted into the flowing LNG stream either directly or via a bypass loop. Laser light is emitted from the end of the probe into the LNG sample, and the scattered Raman light is collected back through the same probe tip. The collected Raman light travels through a second fiber-optic cable and then enters a detector in the analyzer, where the resulting individual wavelengths are identified and quantified. All electronic components are housed inside the analyzer enclosure (**FIG. 5**).

This approach has several critical advantages when compared to GC analysis, including the following:

- The probe inserts directly into the LNG stream so that it takes the reading in situ, with the moving liquid constantly refreshing the sample.
- Measuring LNG in situ means that there is no vaporizer and no sample lines, valves, heaters or regulators. This results in more stable measurements from a Raman analyzer, without the noise and spikes that are characteristic of vaporizer variability.
- The probe can handle pressures up to 111 barg (1,610 psig) for C276 alloy or 87.5 barg (1,270 psig) for a hybrid metal combo, and at temperatures as low as –196°C (–320.8°F) for both material options.
- The laser and detector are housed within the analyzer, and the light is carried to and from the probe via fiber-optic cables over distances up to 500 m. The probe contains only the optical system and is rated for installations in Class 1, Division 1 and Zone 0 hazardous areas.

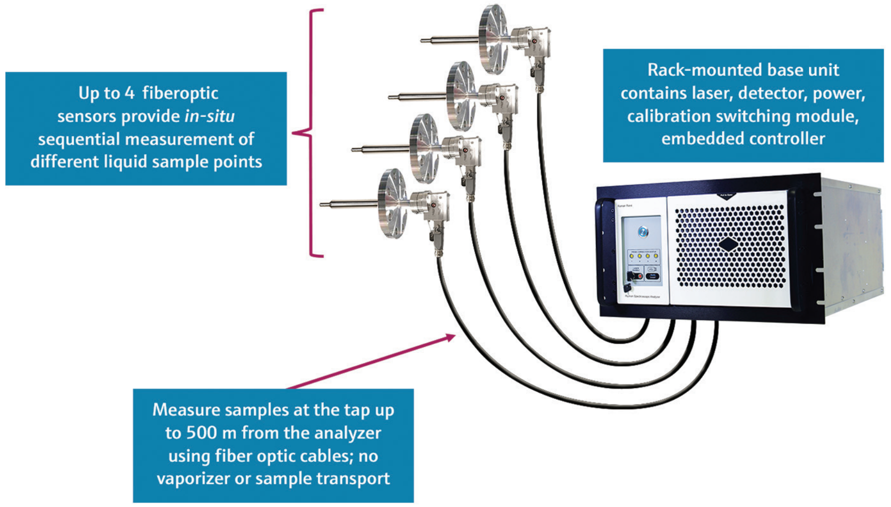


FIG. 4. A proprietary analyzer[b] can take simultaneous readings from up to four probes located in different parts of the process stream.

- A single analyzer can support up to four probes, so readings can be taken at multiple locations in the process stream.
- Output from the probe changes in real time, and the analyzer can take a snapshot of the composition in less than 20 sec, with no delay between readings.

Compositional measurements with Raman spectroscopy can be used to calculate Wobbe index or calorific values, according to International Organization for Standardization (ISO) and Gas Processors Association (GPA) Midstream Association standards such as GPA 2172-09/GPA 2145-2009, GPA 2172-09/GPA 2145-2016, ISO 6976-1995E and ISO 6976-2016.

The Raman analyzer can detect most natural gas components down to 200 ppm, so it is not suited for trace analysis of LNG contaminants—such as H_2S or CO_2—down to regulatory levels. For practicality, LNG pretreatment typically reduces contaminant levels well below specified thresholds. Instead, the calorific value is the more critical measurement, which Raman analyzers can handle within the required ranges. If it is necessary to measure lower contaminant levels, GC analysis is a better solution.

Analyzer calibration. All analyzers require periodic calibration, but the frequency and complexity of the calibration process varies. GC models used for LNG terminal service typically require daily calibration, which is performed by feeding a pre-measured test gas into the analyzer. While this is not normally an arduous process, it requires consumables and taking the analyzer offline. Although calibration can maintain GC measurement accuracy, it does not solve the larger issue with the vaporizer and sampling system. GC can provide highly precise analysis, but if a sample is not representative of the LNG composition, then the underlying problem remains unsolved.

Raman analyzers also require calibration; however, their mechanisms for measurement are very stable, so these analyzers can typically operate up to 2 yr without calibration. Because their probes are inserted directly into an LNG stream, most calibration requires process shutdown for probe removal, which can be performed during planned facility maintenance cycles. Mounting mechanisms are available in certain applications, enabling online probe removal while LNG is flowing.

Analyzer validation can be performed using a surrogate liquid sample. If the validation result indicates that calibration is required, standard calibration tools are available for field calibration of the analyzer to ensure that it is brought back to factory specifications. The calibration procedure tests the entire analyzer, including the probe, fiber-optic cables and the full analyzer.

Importance of analysis. LNG—coming from a variety of sources, each with unique characteristics and worth hundreds of millions of dollars—is moved around the world every day. While pretreatment removes some of the variability, even slight changes in calorific value can change the value of a single load by hundreds of thousands of dollars.

Analyzer technologies available today can ensure that natural gas meets the required quality specifications prior to liquefaction, and then identify when it suffers value loss during extended time in transit. New techniques provide greater measurement accuracy and reliability, combined with lower lifetime costs and maintenance, and receiving terminals should consider all available measurement options. HP

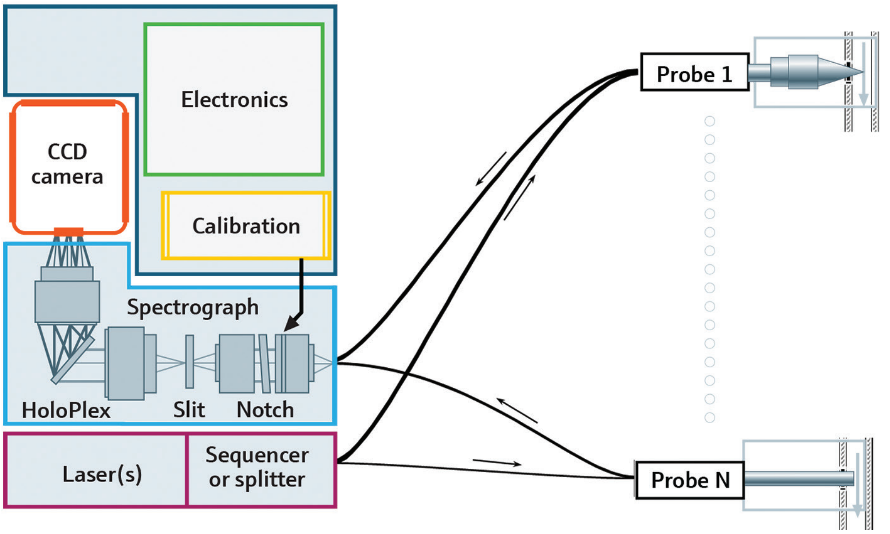


FIG. 5. The ability to insert a probe directly into the LNG stream eliminates the need for a traditional sample handling system.

NOTES

[a] Endress+Hauser's SS2100 TDLAS
[b] Endress+Hauser's Raman Rxn5 analyzer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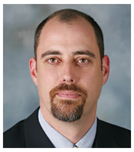
SAM MILLER is the Head of Technical Marketing for TDLAS/QF at Endress+Hauser Optical Analysis. He has more than 20 yr of experience in oil and gas markets and in the development of laser-based products. He is a member of the ASTM D03 standards committee and participates in numerous natural gas conferences, including American Gas Association (AGA) conferences, ISA Analysis Division symposiums, and meetings for various international hydrocarbon measurement organizations. Mr. Miller holds a BS degree from California Polytechnic University, Pomona, and an MBA degree from the University of California, Irvine.


SCOTT SUTHERLAND is the Product Manager at Endress+Hauser Optical Analysis. He is responsible for Raman analyzers, including the Raman Rxn5 analyzer and its applications in the oil and gas and chemical industries, including ammonia and methanol production, hydrogen generation and synthetic natural gas. Dr. Sutherland has spent more than 32 yr designing Raman analyzers and developing Raman solutions for General Electric, Thermo Electron, Photon Systems, Bruker Optics and Endress+Hauser in the areas of law enforcement, forensics, pharmaceuticals, hazardous materials and process gas analysis. He earned BS degrees in physics and chemistry from the University of Tennessee at Chattanooga, and a PhD in analytical chemistry from the University of Florida.

HP Bonus Report

Gas Processing & LNG Industry Spotlight

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Why compact actuation solutions are ideal for constrained space installations

Many oil and gas industry applications face space constraints. This limitation requires more compact solutions providing the same performance as traditional alternatives, and this is particularly true in many valve installations, where actuators often require an inordinate amount of space.

This article will examine two applications—a floating production, storage and offloading (FPSO) vessel, and a floating liquefied natural gas (FLNG) vessel—where space is at a premium, although the problems and solutions discussed are often found in many other areas of the oil and gas value change.

The ultimate space constraint. The landscape of oil and gas production is changing very rapidly. The LNG market has significantly expanded, driven by the soaring cost of energy, as well as by a global desire to switch to lower carbon fuel alternatives.

To reduce costs and make smaller, deepwater oil and gas reserves economically feasible, many producers are turning to offshore processing units. These FPSO vessels are becoming increasingly common all over the world, but their compact design creates significant challenges for engineers tasked with putting so much equipment in tight spaces.

One glance at an FPSO unit tells the story (**FIG. 1**). Unlike a typical onshore oil and gas production facility that has more space, an FPSO facility must jam an entire oil and gas production plant into a very limited footprint.

FPSO vessels are particularly effective in remote or deepwater locations where running seabed pipelines are cost prohibitive. These units also eliminate the need to run costly long-distance pipelines to onshore processing facilities, and they can be moved to another location when a field's production tapers off. However, their main disadvantages are the obvious limitations on available space, along with particularly harsh saltwater environments. Designing all equipment to fit in such close quarters, while allowing this equipment to be maintainable, is a very challenging engineering problem.

The drive toward developing natural gas reserves has spawned an offshore floating facility devoted to natural gas processing, storage and transfer (**FIG. 2**). These vessels (i.e., FLNG units) are like FPSO units, as they accept and process natural gas from a variety of fields, store LNG as required, and then transfer it to large ships for transport around the world.

FLNG production units face the same challenges as FPSO ships in that they must wedge an enormous amount of equipment and piping into a very limited space. However, the FPSO designer's task is made even more difficult by the high pressures, cryogenic temperatures and very large line sizes common in the LNG industry. Fitting 100 automated valves in a small footprint is difficult. Fitting a hundred 24-in., 900-class automated valves in that same constrained space is even more of a vexing problem.

Standard actuators do not fit. A scotch yoke or rack-and-pinion actuator (**FIG. 3**) works well in a facility where there are no space constraints. A typical actuator is mounted above the quarter-turn valve, and it has a pneumatic or hydraulic cylinder on one side of the shaft and an opposing spring cylinder on the other. The pneumatic cylinder uses air pressure to drive the valve to the actuated state, and the spring drives the valve back to the fail-safe state upon loss of air or power. This design is proven and robust, operating reliably in millions of valve applications all around the world.

Unfortunately, typical actuator designs have a very large footprint, especially when the pneumatic cylinder and opposing spring are sized to handle very high pipeline pressures. Even when mounted with the actuator running parallel with the pipe, sufficient space in all directions must be left open to access and replace the actuator, if necessary. In an FPSO, FLNG or other space-constrained environment, this type of actuator is not feasible in some cases because installation space is not available.

A better solution. Fortunately, a new actuator design dramatically reduces the footprint of the actuated valve. Helical-slot actuators combine the pneumatic cylinder and spring cylinders into a single vertical housing mounted directly to the top of the actuated valve (**FIG. 4**).

As the air pressure pushes the cylinder downward, rollers move in the helical slot, converting the vertical movement of the cylinder into rotational movement to turn the valve. When the cylinder pressure is released, the spring drives the cyl-

FIG. 1. An FPSO gas and oil facility is the ultimate study in space management, as every component must be carefully chosen to fit within a very tight envelope.

inder upward, reversing the valve motion. The helix slots can be reversed to provide either fail-open or fail-closed movement. The slot can also be characterized to create different torque curves and valve opening movement profiles to match the valve being actuated.

The resulting space savings can be dramatic, as shown in **FIG. 5.** By combining the pneumatic cylinder and spring into the same housing, the overall actuator size is greatly reduced. More importantly, the actuator is mounted vertically above the valve, allowing the entire valve/actuator assembly to be pulled straight up for easy maintenance.

Since the actuator takes up nearly the same footprint as the valve, this frees up significant space in all directions that can be used by other equipment. In a constrained-space environment, this can make all the difference for the engineering team, enabling this team to fill that space with additional piping. It also enables the team to place valves in very close proximity to each other, or to place them near a wall or tank, which is not possible with traditional actuators. Despite the very close quarters, each valve assembly can be pulled up and out of the pipeline for service and maintenance.

The helical-slot actuator design is an excellent choice for any environment where space is at a premium, such as with piping manifolds and on process skids. The available actuator response speeds and torques match the offerings of scotch yoke or rack-and-pinion actuators, but this functionality is provided in a much smaller footprint. The actuators are also offered as a part of a valve package that is safety integrity level (SIL) rated, enabling them to

FIG. 2. An FLNG facility is as space constrained as an FPSO unit, but it also adds the challenge of fitting in many large, high-pressure, actuated valves—each critical to production and safe operation.

FIG. 3. A typical actuator uses a pneumatic cylinder to actuate the valve and an opposing spring cylinder to drive the valve to a fail-safe state. The design works well, but takes up a great deal of space when sized for large valves and high pressures.

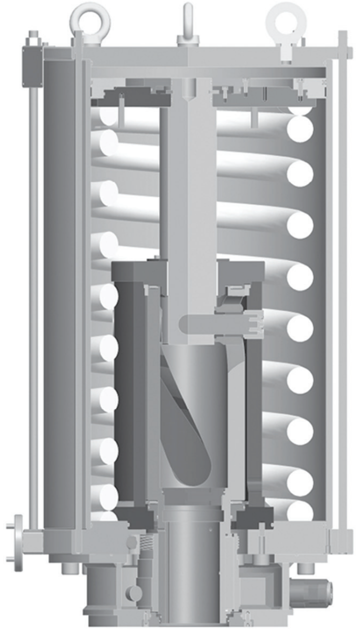

FIG. 4. A helical-slot actuator combines the spring and pneumatic cylinder into a single vertical housing mounted directly above the valve. As the piston moves, rollers engage slots on the side of the helical cylinder to turn the valve.

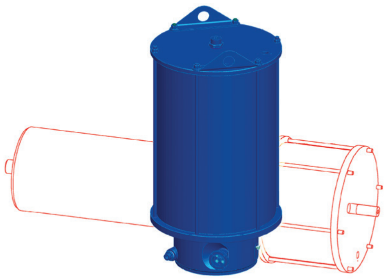

FIG. 5. This picture compares the overall footprint of a scotch yoke actuator (in red) to that of a helical-slot actuator with the same torque. The vertical design of the helical-slot actuator (in blue) allows it to be pulled straight up, requiring little more space than the valve itself.

be used in safety-critical applications, such as in an LNG high-integrity pressure protection system (HIPPS) and with various safety interlock and shutdown services.

Meeting the challenge. An FLNG design firm issued a proposal request for three 24-in., Class 900 safety shutdown valves to meet very demanding requirements for physical size, torque, speed of response and environmental conditions. The actuator torque had to be 1.5 times the maximum torque required for the valve, and since the valves were part of a HIPPS installation, each valve assembly required a SIL 3 rating. The performance specification required the valves to stroke from 100% to 20% in less than 4 sec, and then to close the remaining 20% in an additional 4 sec to avoid damaging the valve seat. The entire assembly had to be painted and coated for an offshore marine environment.

After evaluating the options, actuators utilizing helical-slot designs were chosen due to the dramatically reduced valve footprint. These proprietary helical-slot actuators[a] were paired with SIL-rated digital valve controllers to provide remote health monitoring and advanced diagnostics (**FIG. 6**).

The torque requirements were easily met with the standard actuator design, but the stroke requirements required a modified integral quick exhaust valve that could achieve the initial stroke time yet gradually close the valve as it approached the end of the stroke (**FIG. 7**). A customized quick exhaust valve utilized a damper with two bypass valves to adjust the end of travel performance as necessary to meet the required stroke timing and profile.

The entire valve assembly was certified to a safety rating of SIL 3, and it was painted and coated appropriately to meet the difficult marine environmental conditions. The assembled valves were fully tested and certified during a field acceptance test and are operating in an FLNG production platform in the middle of the Indian Ocean.

Takeaway. The proprietary actuator[a] can be customized to achieve specific torque and movement profiles and paired with high-performance digital valve controllers as part of a highly reliable, safety-rated valve assembly suitable for SIL 3 applications. The actuator speed of response and closing profile can also be altered to meet very demanding actuating specifications, and it can be adjusted to provide a "soft close" final stroke.

When faced with very demanding space constraints in piping manifolds, process skids, or offshore floating oil and gas production facilities, helical-slot actuators are often the best choice. They can match the torque and speed of response of scotch yoke and/or rack-and-pinion actuators, yet their vertical profile offers significant equipment and piping design advantages. This greatly reduced footprint enables the valves to be placed much closer together or positioned near walls or tanks, yet still be accessible for maintenance in either case. HP

NOTE

[a] Biffi TPS helical-slot actuators

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FIG. 6. Three SIL 3-rated, 24-in., Class 900 safety shutdown valves undergo testing for an FLNG safety-critical shutdown application.

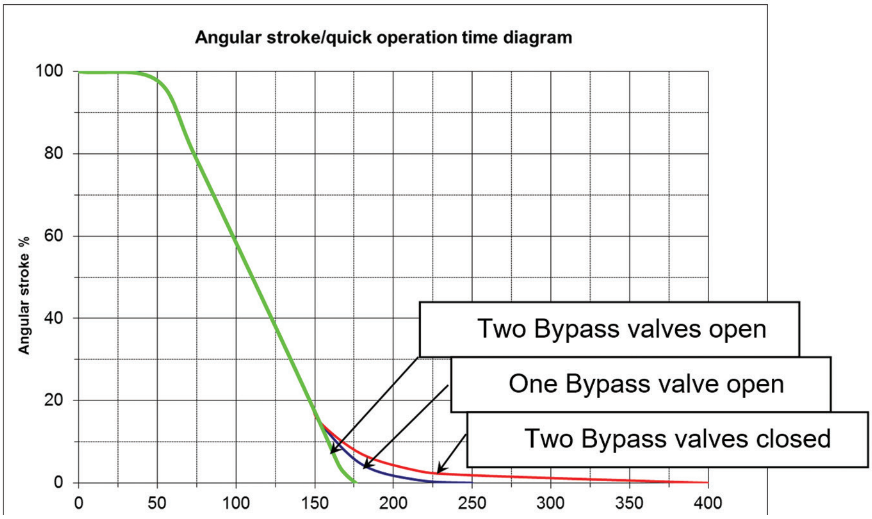

FIG. 7. A customized quick exhaust valve is utilized to quickly close the valve to minimum flow, and to then adjust the speed of the final closure to avoid process pressure surges.

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Soft startup of best practices for the commissioning of amine plants

The startup and commissioning of hydrocarbon plants can present unique challenges that, if not anticipated and addressed in the pre-commissioning and commissioning periods, will inhibit plant startup and increase the difficulty of achieving product specifications in a timely manner. This article provides details on potential multiple challenges during commissioning, as well as a method to avoid these challenges by applying a soft startup concept on a massive scale. The acceptance criteria and successful results of the soft startup concept are detailed here.

Plant A sought to avoid or reduce the common commissioning challenges of a new amine system (e.g., gas treatment facility), including foaming, the presence of unwanted debris or materials in piping and vessels, etc. To ensure the short- and long-term reliability of an amine system facility, Plant A elected to implement a soft startup concept on a massive scale, where the amine circuits are filled completely with pure demineralized water (rather than amine), then pressurized by nitrogen (rather than fuel gas) up to the operating pressure. This was completed for multiple safety and reliability reasons to:

- Test system tightness with a safer medium (nitrogen, as compared to fuel gas)
- Debug the associated process control and emergency shutdown (ESD) systems
- Test the functionality and reliability of all rotating equipment
- Clean the system of any remaining debris and suspended solids prior to charging with expensive solvent into the unit
- Limit the potential for future foaming incidents.

During the soft startup, circulation is started in both cold and hot modes to ensure that all debris and suspended solids are removed from the pipe walls, fittings, valves and other equipment and vessels, and the debris and suspended solids are caught by the strainers and filters. This approach was successfully utilized in the other amine units.

Amine system process description—acid gas removal unit. The gas enters the treating train through the inlet filter separator. The separator is provided to remove any particulate matter or traces of liquids that may be received with the incoming gas. Without a spare separator, a bypass valve (sized for full flow) is installed to allow maintenance servicing of the separator.

The gas coming from the inlet filter is fed to the amine contactor for carbon dioxide (CO_2) removal. The contactor is a tray column divided in two sections. The bottom amine absorber section consists of trays. The lean amine is fed to the top tray and the rich amine is withdrawn

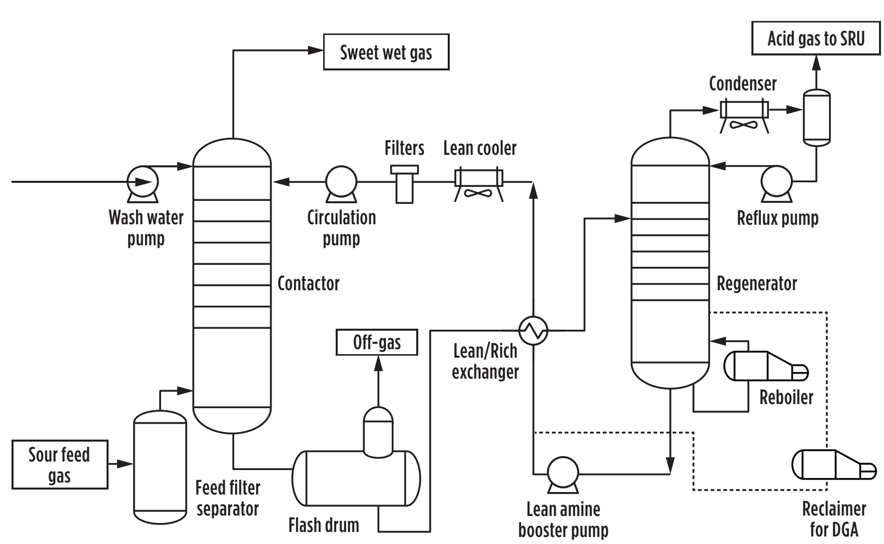

FIG. 1. Typical PFD for an acid gas removal unit.

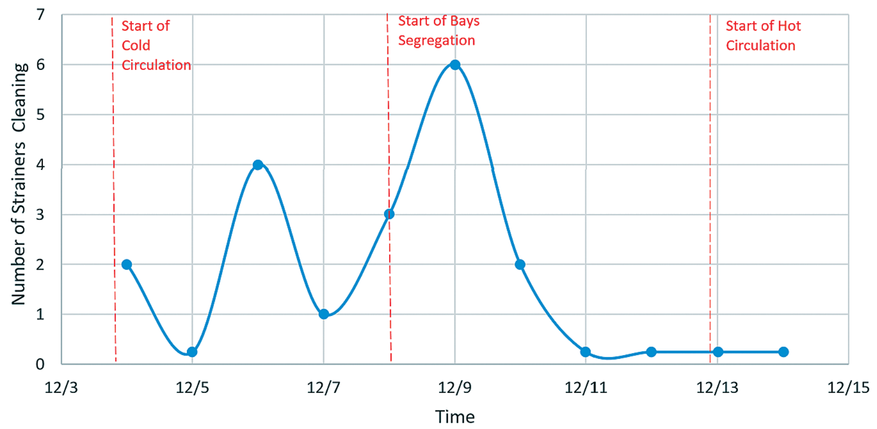

FIG. 2. Frequent strainer cleaning for amine System A.

FIG. 3. The progression of water clarity by the samples collected from amine System A.

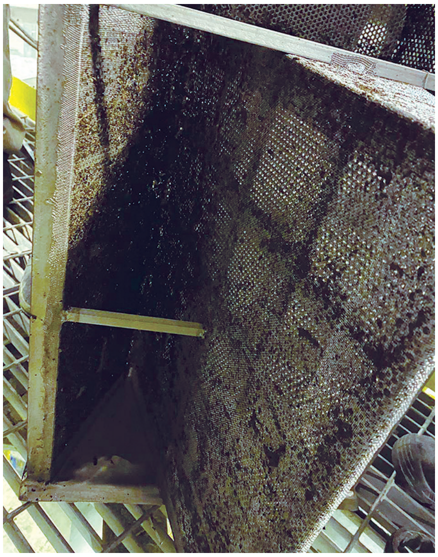

FIG. 4. Dirt accumulation in pump strainers.

from the bottom of the tower. The upper section of the contactor comprises three trays and is used as a water washing system to absorb any amine entrained in the gas from the bottom absorber section. The second pass reverse osmosis water is circulated around the trays by pumps.

A continuous make-up of fresh reverse osmosis water is fed to the top of the section, while excess water is drained from the column water wash chimney tray to the rich amine flash drum. The outlet for the wet sweet gas from the top of the amine contactor is routed out of the gas treating train to the natural gas liquids (NGL) recovery unit.

Rich amine from the contactor bottom is fed to the rich amine flash drum, a horizontal vessel with a top-mounted vertical absorber equipped with random packing. The hydrocarbon gases dissolved in the rich amine are flashed off, washed through the absorber and routed to the thermal oxidizer. As the flashed gases pass through the absorber, any amine entrainment is removed by the counter flow of the water withdrawn from the amine contactor top section. The rich amine coming from the flash drum is heated up through the lean/rich amine exchangers and fed to the amine stripper.

The stripper is a trays column with two once-through exchangers that take feed from the column bottom chimney tray. Heat for the amine stripper reboilers is supplied from the low-pressure hot water system. The gases from the top of the stripper are cooled in the amine stripper overhead condenser and water is condensed in the reflux drum. The acid gas is routed to the CO_2 compression unit. Any organic phase present will be separated in the reflux drum and manually discharged to the wet hydrocarbon burn pit.

The reflux pumps return the condensed water as reflux back to the stripper top tray and, during the regeneration cycle, partially to the amine reclaimer. The lean amine from the stripper bottom is routed toward the amine contactor by booster pumps. Before being fed to the contactor, the lean amine is cooled through the lean/rich amine exchanger and the lean amine cooler. The lean amine circulation pumps boost the pressure and return the lean amine to the amine contactor. Pumps route a fraction of the circulating lean amine (~15%) to the precoat filter package, and filtered amine is rejoined to the lean amine flow.

The amine reclaimer is provided to periodically regenerate the circulating amine. A fraction of the circulating lean amine from the booster pumps (~1% of the total amine pumped flowrate) is fed to the amine reclaimer. The reclaimer operates at 360°F (182°C) and controls the amine solution boiling temperature by an adequate flowrate of reflux water from the amine stripper reflux pumps. Amine and water are recovered to the stripper as vapors coming from the top of the reclaimer, while degraded amine components are left on the vessel bottom. This sludge pond also receives the amine slurry from the amine precoat filters. The acid gas removal process configuration is shown in **FIG. 1.**

Amine system process description—acid gas enrichment unit. The wet acid gas is first cooled in a cooling water chiller and then enters an inlet separator to remove liquid hydrocarbons from the gas stream. The acid gas then enters the bottom of the contactor trayed column and flows upward through the absorber in intimate countercurrent contact with the selective amine solution, where the amine selectively absorbs hydrogen sulfide (H_2S) acid gas constituents from the gas stream. Before leaving the contactor, the CO_2 gas passes through a water wash section.

The rich amine solution is sent from the bottom of the contactor via the rich amine pump to the lean-rich amine heat exchanger, before entering the top of the stripper column. A part of the absorbed acid gases will be flashed from the heated rich solution on the top tray of the stripper. The remainder of the rich solution flows downward through the stripper in countercurrent contact with vapor generated in a steam reboiler. The reboiler vapor (primarily steam) strips the acid gases from the rich solution.

The acid gas and the steam leaving the top of the stripper pass through a condenser, where the major portion of the steam is condensed and cooled in an air cooler. The

acid gas is separated in the reflux drum and sent to the acid gas enrichment unit.

The lean amine solution from the bottom of the stripper column is pumped by a booster pump through a lean-rich heat exchanger and then through air coolers to the top of the contactor. To ensure cleanliness of the circulated selective amine solution in the contactor, a 15% slipstream of the lean solution is passed through a mechanical filtration system.

Acceptance criteria for soft startup. A soft startup can be declared successfully completed when at least two of the following three criteria are achieved:

- The frequency of strainer cleaning is reduced to less than one cleaning activity per strainer every two days.
- Qualitative proof that water samples are getting clearer every operating day for three consecutive days, then stable/no change for an additional day.
- Stable water chemistry parameters are achieved for at least two consecutive days through lab sampling.

PLANT TEST RUNS

Amine System A. Amine System A was placed under a soft startup on December 4, 2019, and in cold circulation until December 13, 2019. During the cold circulation phase, four of the lean amine cooler bays (50%) were isolated and amine was run through the other four bays to increase turbulence and ensure cleanliness (**FIG. 2**). After the system was stabilized and the in-service bays were deemed clean, the bays were isolated and the other four bays were put in service. The switching practice continued until consistent readings on the pump strainers were achieved. Following that, amine System A was placed in hot circulation.

As previously highlighted, one of the acceptance criteria of the soft startup is sample bottle clarity. **FIG. 3** shows the progression of water clarity by the samples collected from amine System A. It can be clearly seen that the water in the system became clearer and cleaner during the soft circulation period.

The first sample (on the right) shows the highest level of dirt. Moving to the left, the samples progressively show an improvement in water quality.

Fin fan tubes were cleaned by limiting flow to four bays at a time, which allowed turbulence flow across the fin fan tubes.

TABLE 1. Lab analysis for water samples, amine System A

	Amine System A			
Analysis	**12/11/19**	**12/12/19**	**1/14/19**	**Unit**
Conductivity	428	430	422	micros/cm
TDS	210	215	211	ppm
TSS	3	3	7	ppm
Total iron	0.98	1.1	1.08	ppm

During this cleaning, a sudden increase in deferential pressure was noticed in all pump strainers due to dirt accumulation, as shown in **FIG. 4.**

Lab analysis of water chemistry parameters (**TABLE 1**) from amine System A showed them becoming stable for at least four consecutive days through lab sampling, indicating that amine System A became cleaner.

TABLE 2. Lab analysis for water samples, amine System B

	Amine System B			
Analysis	**12/11/19**	**12/12/19**	**12/14/19**	**Unit**
Conductivity	718	708	708	micros/cm
TDS	359	354	354	ppm
TSS	5	9	5	ppm
Total iron	0.41	0.47	0.45	ppm

TABLE 3. Lab analysis for water samples, amine System C

	Amine System C			
Analysis	**12/11/19**	**12/12/19**	**12/14/19**	**Unit**
Conductivity	66	74	80	micros/cm
TDS	33	37	40	ppm
TSS	17	2	2	ppm
Total iron	3.14	1.5	1.42	ppm

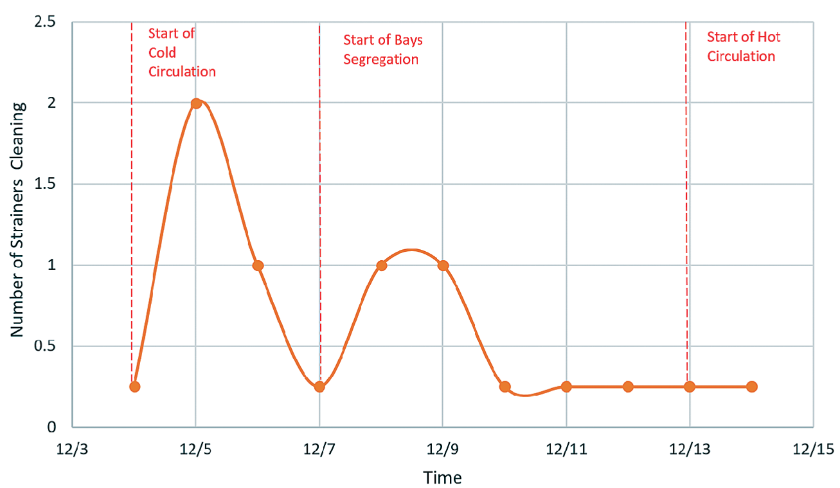

FIG. 5. Cleaning frequency for amine System B.

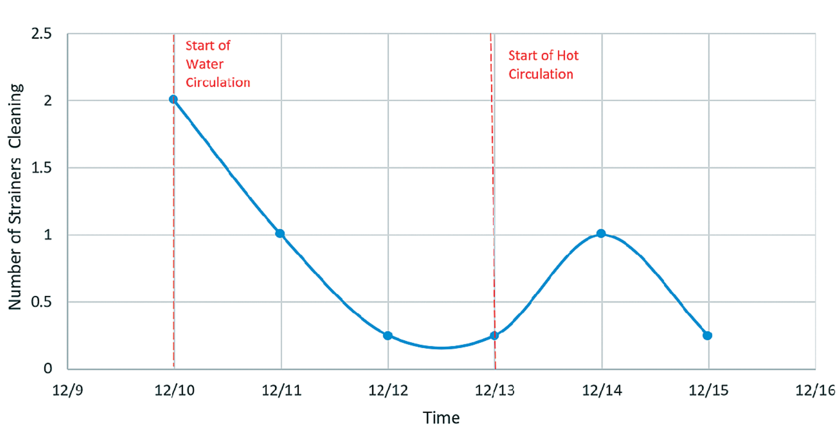

FIG. 6. Frequent strainer cleaning for amine System C.

Amine System B. The system was placed under soft startup on December 4, 2019, and in cold circulation until December 13, 2019. During the cold circulation phase, two of the lean amine cooler bays (50%) were isolated and amine was run through the other two bays to increase turbulence and ensure cleanliness. After the system was stabilized and the in-service bays were deemed clean, the bays were isolated and the other two bays were put in service. The switching practice continued until consistent readings on the pump strainers were achieved. Following that process, amine System B was placed in hot circulation (**FIG. 5**).

Fin fan tubes were cleaned by limiting the flow to two bays at a time, which allowed turbulence flow across the fin fan tubes. During this process—as with System A—a sudden increase in deferential pressure was noticed in all pump strainers due to dirt accumulation.

Lab analysis of water chemistry parameters from amine System B (**TABLE 2**) showed them becoming stable for at least four consecutive days through lab sampling, indicating that amine System B became cleaner.

Amine System C. The system was placed under soft startup on December 10, 2019, and in cold circulation until December 13, 2019. The system was then placed in hot circulation. During the water circulation phase, the amine System C circulation pump strainers were cleaned frequently. The cleaning practice continued until the system become clean.

Lab analysis of water chemistry parameters from amine System C (**TABLE 3**) showed them becoming stable for at least four consecutive days through lab sampling, indicating that amine System C became cleaner.

Takeaway. Based on results from experimental runs on three units, the soft startup concept has proven advantageous, helping avoid common commissioning challenges of a new amine system (e.g., gas treatment facility)—including foaming, the presence of unwanted debris or materials in piping and vessels, etc.—and ensuring the short- and long-term reliability of an amine system facility. HP

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Inspection and test plan: A tool for quality control alignment during construction

Quality control (QC) is fundamental for capital project success. It should be an integral part of how construction is executed. Compliance to project requirements (e.g., drawings, specifications and standards) can be verified through this process. Increasingly, project stakeholders demand robust QC programs to ensure lasting results.

However, practitioners can testify to the complexity of QC processes, which can lead to misalignment among project participants. The owner, general contractor and trade contractors each have a role to play in this process, but getting everyone on board with the process can be a monumental task. Fortunately, there is one tool through which such alignment can be reached: the inspection and test plan (ITP).

What is the ITP? The ITP is a document that contains a detailed list of QC activities that should be performed by the site team from multiple parties in accordance with the defined quality requirements. In other words, the ITP functions as the overall roadmap of QC activities. The inspection and test activities listed in the ITP include the activities before, during and after the construction process.

The importance of the ITP for alignment. Any process that involves multiple parties for implementation runs the risk of misalignment. If this risk eventualizes in the QC arena, the consequences could be serious. When the correct QC requirements are not clearly communicated to the constructor or other parties as to what their roles are in inspection activities, significant rework or, in worst-case scenarios, major safety incidents could ensue because of the violation of design requirements.

The creation of the ITP in a collaborative manner offers a practical approach to bring all parties together to finalize the detailed QC requirements and interfaces. This interdisciplinary process facilitates constructive discussions to determine the suitable level of inspection conducted by the key players.

The agreement on this document provides project leadership the confidence that the right processes are in place to fulfill the quality requirements. Establishing this confidence in a systematic way is the noble goal of any quality assurance (QA) program. In other words, the ITP serves as a bridge between QC and QA programs (**FIG. 1**).

Components of the ITP. ITPs commonly contain the following items:

1. Inspection and test points
2. Acceptance criteria
3. Recording requirements
4. Scope of responsibility.

A sample ITP is shown in **FIG. 2.** The upfront agreement on each of these items has an enormous impact on construction QC efforts.

Inspection and test points. In this section, the project scope is broken down into distinct steps to identify QC activities. The joint review of these items is both a scope and QC alignment opportunity among engaged parties.

This list of QC activities will serve as a valuable checklist to ensure that all necessary inspections and tests are accounted for in the construction process. In most cases, an inspection or test cannot be completed later if it was missed. The ITP helps the team stay on track vis-à-vis all QC steps.

Acceptance criteria. All parties should be on the same page as to what "done" means for construction tasks. The joint review of the ITP could facilitate the agreement on the applicable codes, specifications and other requirements that must be met to complete the

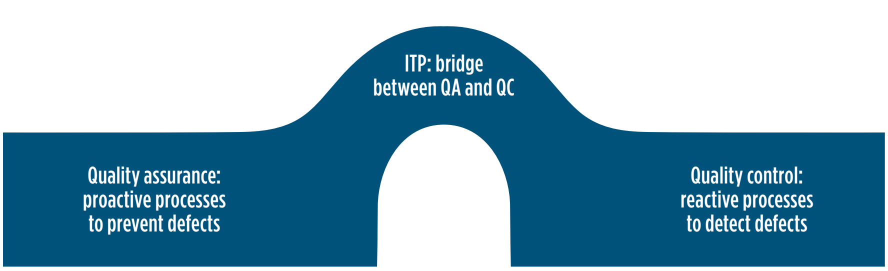

FIG. 1. The ITP functions as a bridge between construction QA and QC.

project's scope. While such references can be found in various places in construction contract documents, bringing them all together in one single document helps with the joint understanding and implementation of them.

In addition, the agreement on acceptance criteria matters when the payment to the contractor is tied to the completion of certain milestones. Having clarity on what constitutes the acceptance of work reduces the likelihood of disagreements on sensitive payment issues during and after construction execution.

Recording requirements. How the results of inspections and tests are documented must be agreed upon upfront to avoid conflicts during QC activities. The retention of test and inspection evidence is critical for historical data capture, legal and/or training purposes in many organizations. Since the owner, general contractor and trade contractors may have different formats for collecting this evidence, addressing what the agreed-upon format is brings efficiency to QC documentation.

Scope of responsibility. What the role of each entity is in the QC activities is a consequential coordination area. The ITP preparation serves as an alignment opportunity to determine the extent of QC activities completed by each party. Examples of QC tasks designated to project participants include to:

- **Inspect:** Physical checking of an activity
- **Witness:** Presence of a representative for a check point
- **Review:** Document reviews to verify completeness and accuracy.

In this section, the parties agree on hold points at which the construction process cannot proceed without the presence of certain parties and their approval.

Translating contractual requirements into the ITP.

The contractor who prepares the ITP should have a good understanding of the QC requirements in the contract. The QC requirements in construction contracts can be found in various places, such as:

- Design specifications
- Drawings
- Local/international standards referenced in the contract
- Manufacturer requirements.

For complex projects, it can be a difficult task to capture all the required checking processes. The preparation of an ITP in a collaborative manner gives the project team an avenue to apprehend QC requirements and to plan on performing them accordingly.

This is also important, considering the prospect of conflicting information in the contract. Such conflicts in the QC arena can come to light during the ITP preparation process and can be resolved by obtaining the owner's concurrence on what requirements take precedence.

Now that the QC tasks are collected into one place during the ITP review and approval process, the owner gets a chance to decide how the QC activities are commensurate with the criticality of the subject construction tasks. This may lead to granting waivers for reducing QC efforts for non-critical items or increasing inspection for more important tasks.

ITP as a dispute avoidance tool.

Unfortunately, the construction industry is a litigious industry and billions of dollars are spent annually for dispute resolution. Many construction claims are originated from quality-related issues. With a close examination, the ITP continues to be an effective instrument to prevent the escalation of such disputes.

For example, the owner's rejection of work due to it not meeting a specification or other requirements that the constructor was not aware of has the potential to be a costly and lengthy litigious matter. Parties to the contract can avoid this by thoroughly reviewing the acceptance criteria noted in the ITP. Alignment on what matters to the owner for accepting the work is important for smooth construction execution and for the prevention of disputes down the road.

Takeaway.

The effectiveness of QC efforts in the construction phase of a capital project requires alignment between involved parties. The ITP is an excellent tool to achieve such alignment. It coordinates project participants on the detailed steps of QC activities, the acceptance criteria, documentation requirements, and the roles of each party in inspections and tests before, during and after the construction processes. The ITP is a medium to gather convoluted QC requirements spread among contractual documents into a single document that can be reviewed and agreed upon collaboratively among project participants. The thorough review and agreement on the details of the ITP can serve as an effective tool to avoid future disputes that could derail project outcomes. HP

Activity No.	Inspection / Test Points	Acceptance Criteria	Recording Form	Scope Of Responsibilities	
				Constructor	Owner
3.0 Welding					
3.2	Confirm the right welding process is being used and equipment is calibrated	Welding Procedure Specification, Weld Map, Project Specifications	Form No. 1234	Inspect	
3.3	Confirm Correct Pre-heat Temperature And Uniformly Around The Weld Joint	Welding Procedure Specification, Weld Map, Project Specifications	Form No. 1234	Inspect	
6.4	Monitor Pipe Temperature During Post-weld Treatment	Welding Procedure Specification, Weld Map, Project Specifications	Form No. 1234	Inspect	
6.5	100% Final Visual Inspections On All Welds	Project Specifications	Form No. 1234	Inspect	
6.6	Non-destructive Examinations (If Required)	Asme Section V. Project Specifications	Form No. 1235	Inspect	Hold Point

FIG. 2. Sample of an ITP.

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Use CFD analysis to evaluate the impact of hot air recirculation on air-cooled heat exchangers

Air-cooled heat exchangers are widely used in the petroleum, petrochemical, natural gas and other industries for cooling purposes. In contrast to a water-cooled system, which consists of a traditional shell-and-tube heat exchanger and a cooling tower, an air-cooled heat exchanger uses ambient air as the main cooling medium. The air is blown from the exterior of the bundle to cool the higher-temperature process fluid inside the tube to the desired operating temperature. Since this procedure does not require a large amount of industrial water, air-cooled heat exchangers are suitable in areas that face geographical restrictions, as well as regions that lack water resources or are required to save water. However, as hot air discharged from an air cooler recirculates, a backflow phenomenon occurs, causing the temperature of the air used as the cooling medium to increase, leading to decreased operational performance of the air cooler.

Several factors play a role in hot air recirculation. Among them, ambient air temperature, external wind direction/speed and the height/distribution of the surrounding buildings are the most important factors. While ambient wind direction and speed are natural forces beyond human control, computational fluid dynamics (CFD) software can be used to conduct flow field analysis and simulate air circulation of air coolers under various environment conditions to avoid adverse effects from hot air recirculation. The most suitable configuration and improved measures can be determined to ensure the heat-exchange efficiency of the whole system meets operational requirements, regardless of wind direction and speed in different seasons.

CFD model formulation. In an overall field area, building location, building height, wall size, direction of entry and exit access, and relative position of each building determine the environmental boundary conditions, which in turn affect the operational performance of the air-cooled heat exchanger. Since hot air moves vertically due to thermal buoyancy and pressure, air discharged from an air-cooled heat exchanger will flow upward and be blown away by the crosswind. Therefore, it is also important to consider the external flow wind direction in the overall impact of boundary conditions.

This article will discuss these aspects based on a refinery engineering, procurement and construction (EPC) project in Southeast Asia. As shown in **FIG. 1,** two major buildings (Side Building A and Side Building B) are on the west side of the main building where an air-cooled heat exchanger is located, as well as a few smaller buildings. A preliminary flow field result suggests that these two major buildings have a greater impact on the flow field than the other smaller buildings because they have solid walls that obstruct air from flowing freely. Therefore, the smaller buildings can be ignored due to their negligible effect on flow field—the focus should instead be placed on simulating the flow field of these three buildings. This also has the benefit of lowering resource loading during simulation. Moreover, all irrelevant factors that may affect rapid convergence of the overall simulation are deleted from the analysis model.

Reference index: Flow field (velocity field and temperature field). Once the boundary condition of the steady flow field is set and the turbulence module calculation is complete, the flow field can be observed through simulation.

To quantify the results of CFD numerical analysis, the now-known flow field phenomenon was combined with various energy condition inputs to produce a temperature distribution map around the

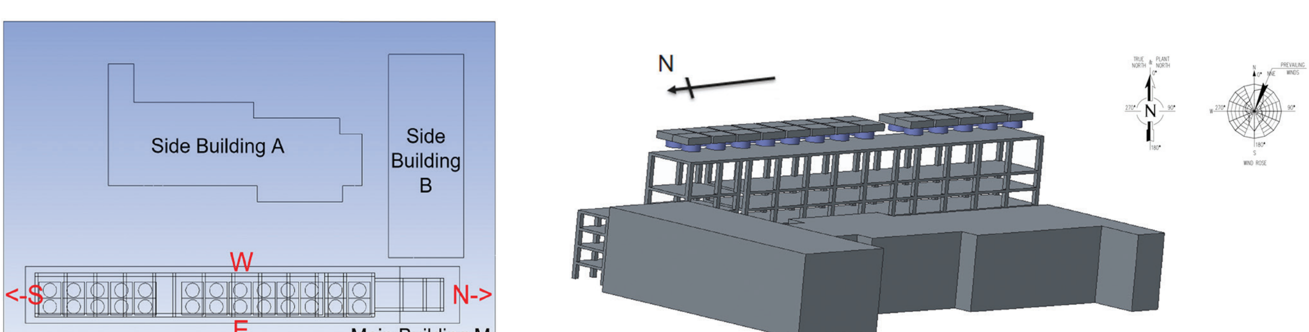

FIG. 1. Schematic plant layout.

building, as shown in **FIG. 2**. Comparing the gradient of the temperature field and flow field diagram shows that temperatures rise more significantly where the flow field is disturbed and where turbulences and vortices occur. In particular, temperatures rise significantly in places where accumulated heat is unable to disperse easily due to slower average flow velocity.

Verification by CFD simulation shows that the air-cooled heat exchanger in the design condition may not produce the ideal outcome in terms of heat exchanging performance. However, further assessment based on different design conditions may be needed.

Simulation result 1: Impact from the angle between the equipment setting direction and crosswind direction. The construction site's wind rose diagram (showing the percentage distribution of local wind direction/wind speed) indicates that the prevailing winds are north-northeast (NNE), north (N) and south (S). Although the north wind, which is the second strongest wind, has little effect on the outflow air volume of the air cooler, the streamline diagram in **FIG. 3** shows that air coming out of the air cooler outlet is clearly affected by the crosswind, as the air discharged from the upstream air outlet is sucked by the downstream fan, causing hot air recirculation. This phenomenon is absent in the case of the strongest wind, the north-northeast wind. The takeaway from this analysis is to avoid displaying air coolers in parallel to the prevailing wind direction to reduce the chance of hot air recirculation.

Simulation result 2: Impact from wind speeds. How hot air recirculated at various crosswind wind speeds when blown by the north wind was also calculated. **FIG. 4** illustrates the changes in temperature as affected by wind speeds. At low wind speed, temperatures at the inlet/outlet accesses rise gradually. As wind speed increases, the trend of temperature rise is reversed and tends to remain stable. Conversely, in terms of the air-cooled heat exchanger's accumulated heat-dissipation performance, small fluctuations were seen within a certain range (**FIG. 5**).

The faster the ambient crosswind, the higher the temperature at the air cooler outlet/inlet (the air cooler outlet temperature peaks at a wind speed of 7 m/sec). However, the temperature begins to drop as the wind speed exceeds 7 m/sec, until finally beginning to fluctuate at 15 m/sec and heading toward dynamic balance.

In a conventional CFD analysis report on air coolers/heaters, a system is considered free from hot air recirculation and, therefore, efficient as long as the temperature at the air inlet is close or equal to the ambient temperature. The challenge, however, is that the design wind speed is generally higher than the actual ambient wind speed and does not consider lower wind speeds. From the simulation results, it was determined that the worst case in terms of higher temperature actually occurs in a low wind speed scenario. Therefore, traditional verification methods may need to be re-examined in this case.

A further examination of the relationship between crosswind speed and an air cooler's heat dissipation capability reveals that the trend of average heat dissipation does not necessarily drop or increase as the crosswind speed increases. This means that the heat dissipation capability does not necessarily decrease as wind increases. This may be due to the fact that when the wind is strong, a large proportion of the heat will be dissipated from the wall of the air cooler, causing less accumulated heat and more overall heat dissipation.

Simulation result 3: Impact from the interval length between bays. In a windless flow field, the air exiting the air cooler outlet should flow regularly upward. Observing the flow vector from a structure elevation view (**FIG. 6**) indicates that air emitted from the outlet fans becomes diffused. Closer to the exit, the outlet air speed is faster and the cross-sectional area is smaller. Conversely, the farther away from the exit, the outlet air speed becomes slower and the cross-sectional area becomes larger. When the distance between two fans becomes too close, the scattered air flow lines are affected, causing turbulent flow. As the crosswind passes, it will interrupt the original flow field between the bays, causing inter-

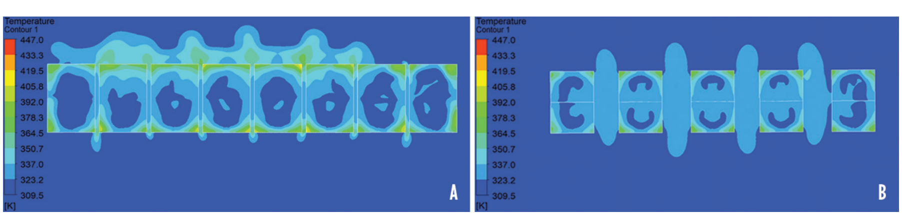

FIG. 2. The temperature gradient of air inlet elevation with various interval lengths between bays: plan view (control group) (A); and the temperature gradient of air inlet elevation with various interval lengths between bays: plan view (experimental group) (B).

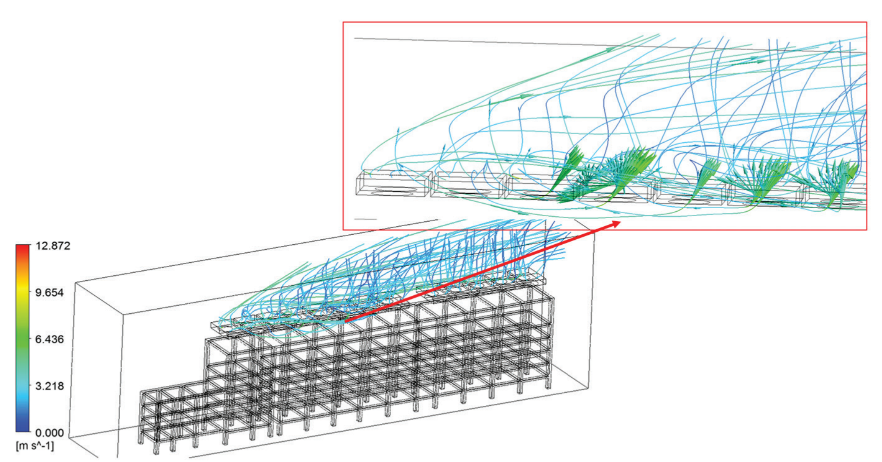

FIG. 3. Streamline diagram showing the effects of north wind.

ference against each other and a resulting change in temperature distribution.

A further comparison can examine how spacing between adjacent bays of the air cooler would impact air flow and heat dissipation by looking at the differences between a control group and an experiment group. Despite the flow vector diagram, in a control group (**FIG. 6A**), the air coming from the outlet of the air cooler flows upward regularly, and adjacent bays in the central area have a greater impact on each other in terms of air flow, as outlet air is obstructed due to the adjacent bays' outlet air pressure. When the temperature plan view of the air inlet is observed, it can be seen that in the control group (**FIG. 2A**), air discharged from the upstream air outlet is sucked by the downstream fan. Such hot air recirculation causes the temperature superposition effect. Coupled with air obstruction at the air outlet due to outlet air pressure from adjacent bays, heat is accumulated.

In contrast, the experimental group with larger interval lengths between bays (**FIG. 6B**) shows a more uniform upward flow of outlet air from the air cooler, as well as a lower mutual impact between bays. The outlet air speed increases, while hot air recirculation becomes less frequent. The inlet temperature tends to be consistent, while the temperature superposition effect is gentle (**FIG. 2B**). With less obstruction at the outlet from the adjacent bays' outlet air pressures, more heat can be removed.

Flow field simulation shows that the impact of spacing between two adjacent bays on air outlet and heat dissipation will vary nonlinearly with interval lengths. Therefore, the optimal air cooler spacing can be determined through tests and simulations. The design of air coolers is often such that bays stand next to each other—when the angle between the display direction of the air coolers and the ambient wind direction sits at 45°, the largest interval lengths between bays are available. Echoing conclusions from a previous assessment on the angle between equipment setting direction and crosswind direction, the best angle should indeed be 45°.

Takeaway. The most effective approach to reduce the impact from hot air recirculation on air cooler performance is to avoid placing the air cooler in parallel to the prevailing wind direction. The author's company's analysis and simulation indicate that a larger temperature rise ratio of the air cooler outlet/inlet occurs at lower wind speeds rather than at higher wind speeds. The analysis and simulation also show that, contrary to the assumption on the heat dissipation efficiency of air coolers, increased ambient wind speeds do not necessarily mean worse heat dissipation capacities. The efficiency depends on the overall cumulative effect of the heat conduction and heat convection. HP

ACKNOWLEDGEMENT

Research results discussed in this publication come from industry-academia cooperation between CTCI Corp. and the National Taiwan University of Science and Technology.

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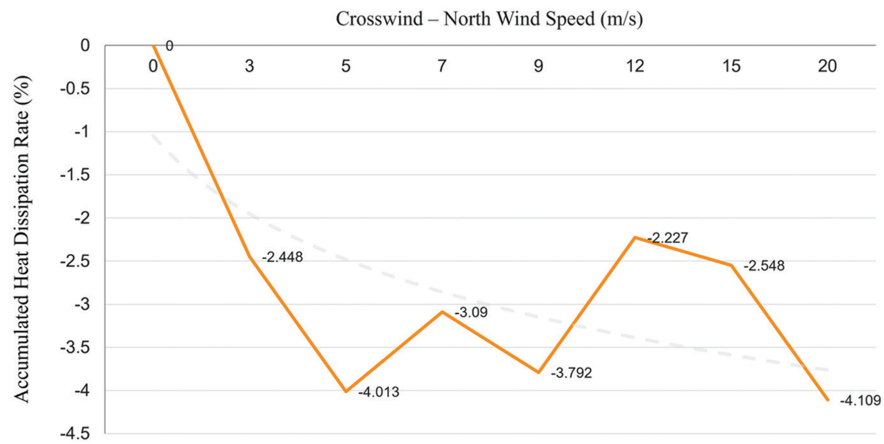

FIG. 4. Temperature rise ratios of air-cooled heat exchanger inlet/outlet in various crosswind speed scenarios.

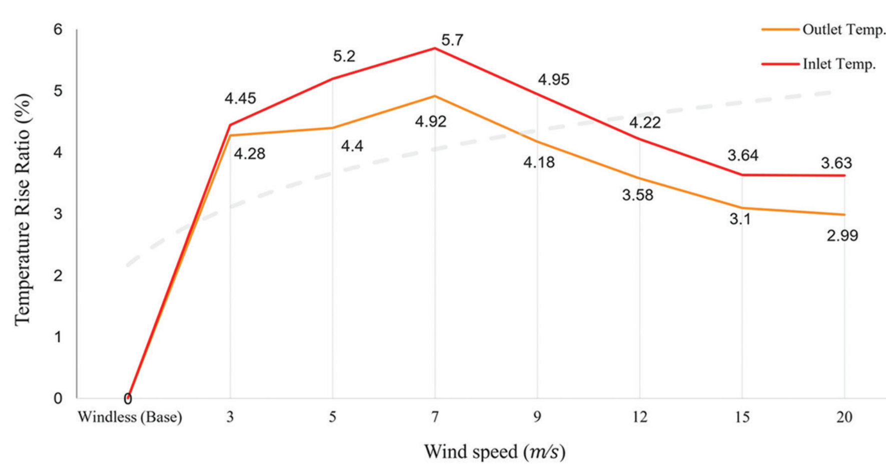

FIG. 5. Accumulated heat dissipation rate of air-cooled heat exchanger in various crosswind speed scenarios.

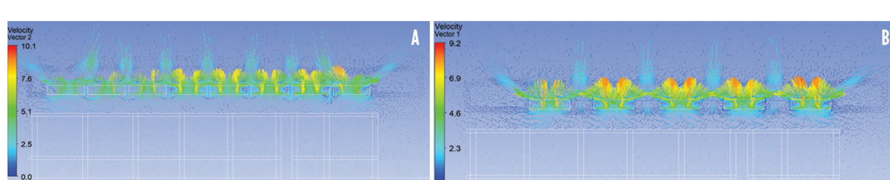

FIG. 6. A flow vector diagram with various interval lengths between bays: elevation view (control group) (A); and a low vector diagram with various interval lengths between bays: elevation view (experimental group) (B).

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Creating opportunities for future fluid machinery experts

The state of understanding science and facts on one side and distinguishing these from anecdotes and often far-fetched opinions on the other side should be concerning to us. Institutions of higher learning can be disinclined to teach needed, applicable or actionable knowledge. While there have been discussions about apprenticeship training, we have also heard selfish comments to the contrary. Trained technical professionals feared that they would be hired away by companies that themselves do not invest in training. As a result of this prevailing mindset, little or nothing is done. Moreover, overdependence on the Internet causes some not to learn how to connect the dots and too often miss seeing matters in the proper context.

In need of more lubrication knowledge. Decades ago, it became evident that lubrication issues in process industries caused disproportionate financial underperformance. Experts knew that the industry could capture large benefits from reliability improvements with a little more forethought. At that time, an engineer approached one of the department heads at a major university and suggested they should add an undergraduate course in lubrication technology.

FIG. 1. Pump craftsmen at work in Cressier, Switzerland.

The department head's answer was swift. He said his university would become a laughingstock for teaching how to use an oil can or a grease gun to refill bearing housings, and he would neither support nor even mention such a recommendation to his fellow educators. In the roughly 30 yr since floating his recommendation, the engineer dealt with ever-increasing lubrication matters. He had ample opportunities to contribute to books on lubrication technology and testify as an expert witness on lubrication-related mistakes in U.S. courts and International Courts of Arbitration in Washington, DC and London, UK. He reaffirmed that lubrication knowledge adds more value than the credit it receives.

Consider imitating the success of others. We have lived long enough to see how reluctant people in authority often are to allow others to comment on matters before deciding on a course of action that may (or may not) be the best. At times, we could learn much from the success of others. In many European countries where craft training is part of the overall educational system, apprentice electricians, mechanics or millwrights must spend 20% of their working hours as apprentices in trade schools. This is typically 8 hr/wk, usually mandated by a country's Education Ministry. This schooling covers the relevant theories and sciences underlying their respective crafts. While the leading U.S. engineering colleges often prescribe curricula for mechanical engineering studies, including a hands-on machine shop course, there is now also a need for trained technicians. The 20% rule or creating slots in community colleges, combined with internships, are suggested solutions. In short, there does not seem to be any downside to simply imitating one of the approaches taken by other successful industrial societies.

A recipe for success. As an example of a recipe for success, we address the possibilities for the U.S. Gulf Coast. Perhaps, this would point to a university located somewhere between Brownsville, Texas and Mobile, Alabama; the institution might consider offering a Bachelor of Science degree in fluid machinery engineering. Top-tier companies that repair and upgrade process pumps, dynamic and positive displacement gas compressors and various types of turbines for customers worldwide would arrange internships. The companies that would make room for these internships might make it one of its priorities to point out the best available equipment-related components and maintenance avoidance strategies. These would include amendments to the American Petroleum Institute equipment standards and foundation and piping recommendations practiced by a few true best-in-class fluid machinery user companies. Interns would learn that customers not ready to authorize certain feasible upgrades on a particular repair would be hard-pressed not to authorize these on one of the repair events that will likely follow (**FIG. 1**).

As to the interns who receive a balanced education in both theory and practice, many would opt to pursue a 2.5-yr program leading to an AS degree in fluid machinery technology. All graduates would likely be hired by fluid machinery user companies with plenty of repeat repairs or where upgrades would capture significantly improved energy efficiencies. As for the failure frequencies of process pumps, the industry desperately needs the kind of talent that feels offended when pumps made up of fewer than 100 components have more failures per lifecycle than aircraft jet engines with perhaps 8,000 parts. The moral is that the industry needs motivated value adders to be productive and profitable in the years ahead.

Institutions for adult education. Since 1982, some readers have become well acquainted with the adult learning and industry training departments of a few universities located along the Texas and Louisiana Gulf Coasts. Short courses and lectures filled an auditorium in Beaumont, Texas 35 yr ago. However, the last such short course (in 2018) captured a mere eight attendees in Lake Charles, Louisiana. The need for solid training has not decreased; however, what interfered is the erroneous notion that everything we need to know is available on the Internet. Frankly, the Internet is a collection of dots or islands of knowledge. The challenge is in connecting the dots, not collecting them. Dots are often the equivalent of out-of-context information, and the last thing an industrial enterprise will ever need is anecdotal or out-of-context information.

Urgent call for motivated subject matter experts (SMEs). Since around 2000, there is considerable evidence that the petrochemical industry has lost thousands of SMEs to attrition. The repercussions are foreseeable, and an actual example will make the point without much drama or unneeded tactful eloquence.

A large petrochemical complex in the U.S. used oil mist lubrication on its electric motors to great advantage for many decades. Years later, and with all the experts presumably gone, a new project was commissioned, but its many electric motors were grease lubricated. Grease lubrication is labor-intensive, and over-greasing or over-pressuring will negatively affect both bearing reliability and maintenance budgets. The motors for the new project were probably bought from the lowest bidder, and the winning bidder likely underbid others by not using the required irradiation cross-linked insulating tape for the T-leads in the mo-

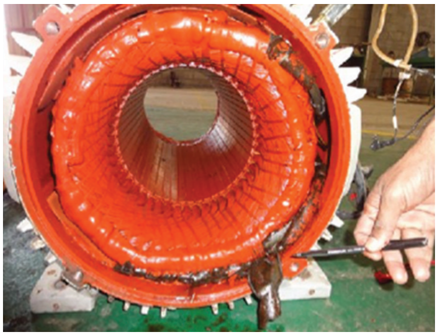

FIG. 2. Swelling standard electrical tape impeded air flow in this motor.

tor. Irradiated cross-linked tape is needed to ensure low swelling rates in the presence of oils or greases. If adhesive-coated insulating tape—as used by electricians in residential work—with its unavoidably high swell rates were being used, the tape would become gummy and unravel when hot (**FIG. 2**). This would impede the flow of cooling air.

Top motor suppliers recognize the responsibility as suppliers of coaxial connectors, cable assemblies and components used within products that are targeted for contact with lubricants. Likewise, reliability-focused user companies make sound choices. Their SMEs will steer an organization towards working with responsible and trustworthy electric motor makers. These SMEs' motor specifications or procurement documents would contain the words "irradiated cross-linked polymer tape" and spell out maximum allowable swell rates in numeric terms.

Manufacturers deserving responsible and trustworthy personnel make it their goal to understand the purchaser's priorities completely. By disclosing the range of capabilities of their products and components, the vendor/manufacturer becomes the purchaser's technology resource. Alternatively, and preferably, a reliability-focused user-purchaser will groom and nurture SMEs whose tasks include the development of rigorous motor specifications. These specifications should ensure the motors in the plants incorporate only confirmed suitable T-lead insulation. Common sense tells us that reliability engineering includes hundreds of details, as highlighted here.

Although a common product will suffice in most applications, standard stretchable insulating tape will not serve well as an oil-resistant cable termination. Oil resistance can also be an important consideration in the terminal boxes of oil mist-lubricated electric motors. Common electrical tape swelling near the stator windings in an electric motor may impede motor cooling effectiveness, and insufficient heat removal can then drastically reduce winding life.

The user company described here has gone back to grease lubrication on all its motor bearings throughout the facility, and its pump and motor mean times between repairs have now fallen well below the values for which the company received much acclaim in 1984. We would recommend they look for future degreed fluid machinery engineers or graduates of a fluid machinery technology institute, which today exists only in our imagination. HP

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Pressure swing adsorption vessels fatigue analysis

The hydrogen (H_2) plant in a refinery produces high-purity (99.9+ vol%) H_2 products using a licensed pressure swing adsorption (PSA) purification process to fulfill the H_2 makeup needs of refinery hydroprocessing facilities, as shown in **FIG. 1.** The H_2 recovery/purification process consists of two trains. The PSA process purifies H_2 by filtering heavier gases/impurities from the process stream as it passes through an adsorber vessel. Each train has 12 identical adsorbers in four groups of three vessels. Each vessel goes through four stages in a cycle: adsorption, depressurization, desorption (regeneration) and repressurization. Adsorption at high pressure pulls all the heavier gases/impurities from the feed gas, producing 99.9% pure H_2. The vessels can handle 40,000 cycles/yr of pressure/de-pressure between 50 psig and 335 psig. This fluctuating pressure range allows for a fatigue design life of 20 yr.

Problem statement. The original equipment manufacturer (OEM) design completed during the manufacturing of the vessels was based on expectations of the pressure cycling range being between 5 psi and 340 psi. The operating cycle time was anticipated to be 12.08 min, which would result in 870,000 cycles in 20 yr. The fatigue design life was acceptable with negligible cumulative usage factors on almost all sections of the vessel except at the joint located between the skirt, and the bottom head was at 87.6%. As a result, the adsorber vessels were designated at 20 yr of design fatigue life with a safety margin of 12.4% at the previously mentioned location, all based on the nominal thicknesses of the vessel components. Since the PSA vessels were nearing their end of fatigue-designed life based on the OEM design, a complete replacement of 24 vessels was required as they were approaching 20 yr in service. However, since the H_2 plant has been in operation, the PSA vessels have experienced changed operating conditions, wherein the actual pressure cycling range was reduced to between 16 psig to 285 psig, as shown in **FIG. 2.**

For an accurate and representative remaining life assessment of the PSA vessel, the thickness of the vessel components was available, and the presence of objectionable flaws had been detected and geometrically characterized by non-destructive examination (NDE)-based inspections. The inspections confirmed that all 24 inspected vessels showed good health, apart from two vessels with linear indications on the shell parent plate and the circular and longitudinal weld seams. These factors established the need for a fitness-for-service (FFS) assessment. The estimated OEM fatigue design life calculations were deemed completed during the manufacturing of the vessels as an unacceptable technical basis for the forthcoming replacement strategy of PSA vessels.

A stress-fatigue evaluation in accordance with the American Society of Mechanical Engineers (ASME) Sec-VIII[3] Div-2/American Petroleum Institute (API)-579[1] Annex B1.52 was required for the components (head, nozzle, flanges, shell and skirt) of the PSA vessels in pressure cyclical service to evaluate protection against failure from cyclical loading. In the assessment, embedded flaws were identified in the vessel wall during the inspection findings and classified as crack-like flaws. In addition to the stress-fatigue analysis, the API-579[1] FFS assessment was required to establish the behavior of the embedded flaws detected during the inspections of the vessel components. Based on the estimated fracture toughness properties of the vessel shell, this analysis would predict the future crack growth and further validate the life extension of the vessel.

Finite element modeling. The finite element method (FEM) with continuum elements was utilized in this analysis, which provided the total stress distribution in the vessel components. The PSA vessel was modeled as a 30° 3D-solid

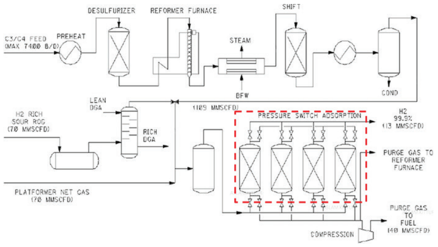

FIG. 1. Pressure swing adsorption in a H_2 plant.

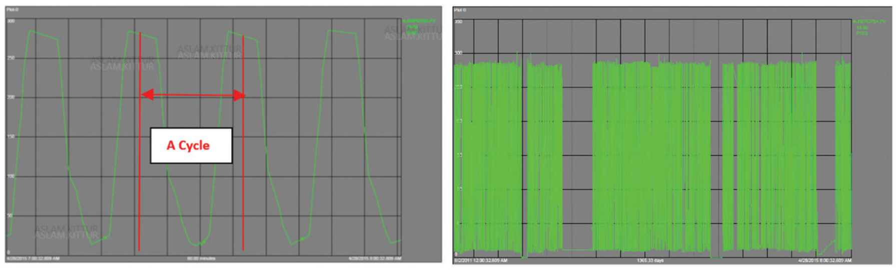

FIG. 2. Operating pressure trends of PSA.

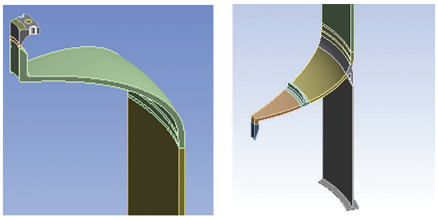

FIG. 3. 30° axis-symmetric model.

model based on the cyclical symmetry, as shown in **FIG. 3.**

This model enables the accommodation of the geometric details responsible for the predicted failure due to fatigue. A varying finite element mesh size ranging from 0.1 in.–0.25 in. was chosen appropriately to accommodate a minimum of five elements in the thickness direction of each pressure vessel component, as shown in **FIGS. 5–10.** The properties shown in **TABLE 1** were obtained from ASME Sec-II Part-D2 and were utilized as input for the analysis.

For carbon steels with C ≤ 0.30% (all the components in **TABLE 1**), the Modulus of Elasticity E value of 29.31 × 10^6 psi at a temperature of 105°F, and the Modulus of Elasticity E value of 29.5 × 10^6 psi at a temperature of 70°F were utilized. The applied loads were appropriately factored down to a 30° (1⁄12th of full load) sector model basis. Symmetric boundary conditions were applied on the entire longitudinally cut edges of the 30° sector model. The loads per **FIG. 4** were applied for the stress and finite element analysis (FEA) model.

The secondary stresses due to general thermal effects were not considered since it was assumed that there were no significant thermal gradients because the operating temperature for the adsorber vessel was 105°F. Stresses and strains produced by any load or thermal condition that does not vary during the cycle were not considered in a fatigue analysis, as the fatigue curves utilized in the evaluation were adjusted for mean stresses and strains and the degree of conservatism employed.

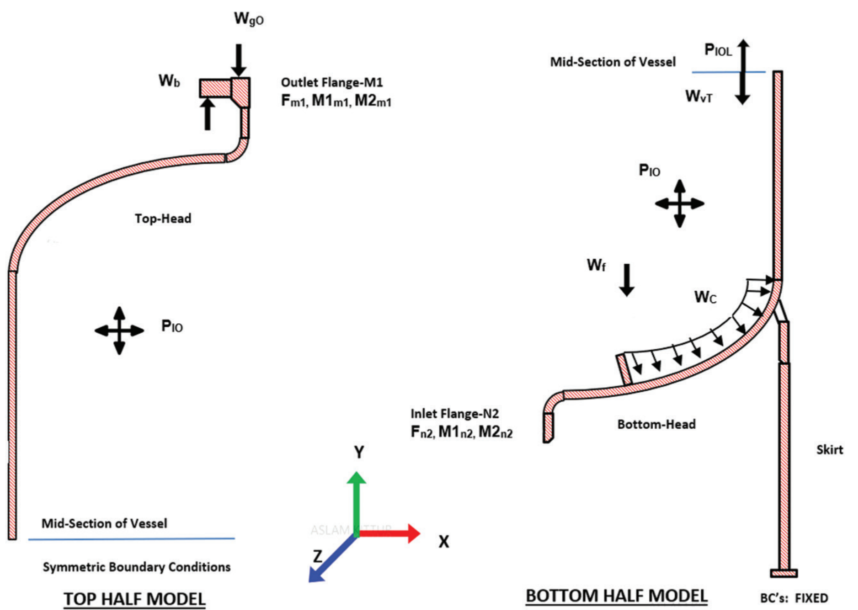

FIG. 4. Loading schematic for FEA model.

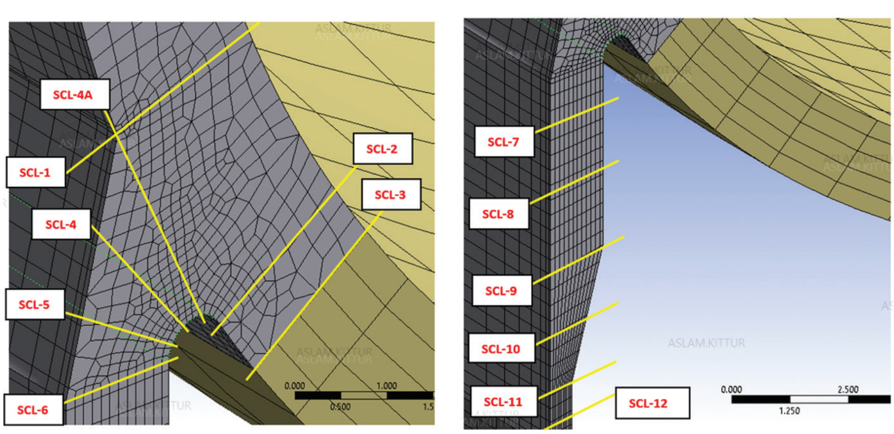

FIG. 5. SCLs in finite element model at skirt weld.

TABLE 1. Material properties (ASME Sec-II Part-D)

Component	ASTM Material Specification	Min. tensile	Min. yield	Max. allowable stress (Sm, Table-2A Sec-VIII, Div.-II)
Shell and heads	SA-516 Gr 70 UNS K02700	70 ksi	38 ksi	23.3 ksi (up to 199°F)
Nozzle-stub, flanges	SA-105 UNS K03504	70 ksi	36 ksi	23.3 ksi (up to 199°F)
Skirt support	SA-285 Gr C UNS K02801	55 ksi	30 ksi	18.3 ksi (up to 200°F)

Analysis methodology. A fatigue assessment was carried out using the elastic stress analysis and equivalent stresses method. The FEM with continuum elements provided the total stress distribution, which was linearized on a stress component basis (membrane and bending stresses categories) using the stress-integration method, and the equivalent stresses were calculated using the maximum distortion energy (von Mises) yield criteria. The effective total equivalent stress amplitude was utilized to evaluate the fatigue damage.

The evaluation for fatigue was made based on the number of applied cycles of the equivalent stress range at every point in the vessel's component. The allowable number of cycles was then evaluated to determine adequacy for the duration of operation to ascertain the suitability for continued operation. The assessment of the embedded flaws in the vessel shell was based on linear stress analysis, where the numerical model does not include the crack explicitly. The acceptance criteria were based on a two-parameter failure

TABLE 2. Loading values

Loads (based on 360° model) for Operating Case:	Symbol	Case-I	Case-II
Internal Operating Pressure	P_{IO}	285 psig	16 psig
Longitudinal pressure (Boundary stress) @ circumferential cross-sectional area of shell (At mid-section)	P_{IOL}	11,849 psi	665.9 psi.
Gasket Reaction Load. The gasket reaction varies with the operating load cycle.	W_{gO}	154,933 lbs	300,439 lbs
Axial force P, inwards along global-X on Nozzle N2 (Gas Inlet Nozzle on Bottom Head) (Non-cyclic)	F_{n2}	22,400 N	
Moment M1 on Nozzle N2 gas Inlet nozzle on Bottom Head (Non-cyclic)	$M1_{n2}$	17,185 lbf-ft.	
Moment M2 on Nozzle N2 Gas Inlet Nozzle on Bottom Head (Non-cyclic)	$M2_{n2}$	13,217 lbf-ft.	
Axial force P, inwards Along global-Y on Nozzle M1 Manway / Gas Outlet Nozzle on Top Head (Non-cyclic)	F_{m1}	7,554 lbf	
Moment M1 on Nozzle M1 Manway / Gas Outlet Nozzle on Top Head (Non-cyclic)	$M1_{m1}$	463,956 lbf-in	
Moment M2 on Nozzle M1 Manway / Gas Outlet Nozzle on Top Head (Non-cyclic)	$M2_{m1}$	356,856 lbf-in	
Operating Fluid Load (Non-cyclic)	W_f	1,200 lbs.	
Catalyst Weight (Non-cyclic)	W_c	92,323 Lbs	
Vessel Weight – Top half (Non-cyclic)	W_{vT}	53,930 lbs	
Flange tightening loads on Ø24" #150, WNRF, Nozzle M1 Manway / Gas Outlet Nozzle on Top Head (Non-cyclic). This bolt load remains constant throughout the operating pressure cycle. (Non-cyclic)	W_b	309,094 lbs	

assessment diagram (FAD) approach to evaluate the combined effects of fracture and plastic collapse per API-579[1] Part 9. The stress values were post-processed to tie into existing K_I and σ_{ref} solutions, as the geometry of the embedded flaws were approximated by simple shapes addressed in Annex C and Annex D of API-579[1]. The crack growth model based on the Paris equation was employed to estimate the remaining life of the vessels with a crack-like flaw based on the linear elastic fracture mechanics approach, where each increment of crack extension shall correlate to a certain increment of stress cycles.

FIGS. 5–10 show the FEM model and the quality of the mesh on the stress classification planes and the location of the stress classification lines (SCLs) selected at the critical locations of gross and local structural discontinuities for this analysis. The stress components will be integrated along the SCLs through the wall thickness to compute the equivalent linearized membrane, bending stresses and peak stress components.

The cases for analyses performed have been developed based on the objective of assessment and methodology for the two scenarios encountered during operation. The stress analysis was performed for the below scenarios:

- CASE I: OPE-MAX internal pressure of 285 psig
- CASE II: OPE-MIN internal pressure of 16 psig.

The stress results from Cases I and II were combined to determine the

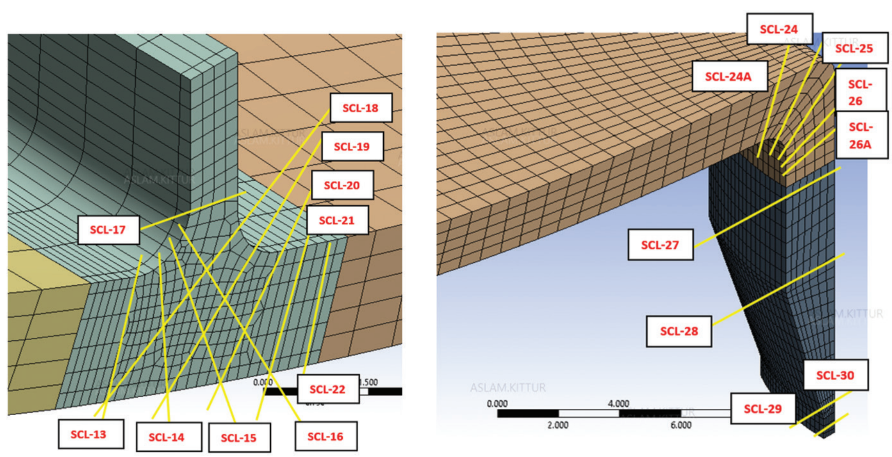

FIG. 6. SCLs in finite element model at bottom head, local.

$\Delta S_{n,k}$ (primary + secondary, equivalent stress range) and $\Delta S_{p,k}$ (primary + secondary + peak, equivalent stress range). The fatigue analysis was completed for the scenario that the PSA vessels of both trains have witnessed 480,000 pressure cycles over the 15 yr in operation based on the available operating data for the uptime of the H_2 plant.

RESULTS–ELASTIC STRESS ANALYSIS

The equivalent stress consisting of primary + secondary + peak was derived from the highest value across the thickness of the section (head, nozzle, flanges, shell and skirt) and was produced by specified operating pressures and other mechanical loads, including the effects of gross and local structural discontinui-

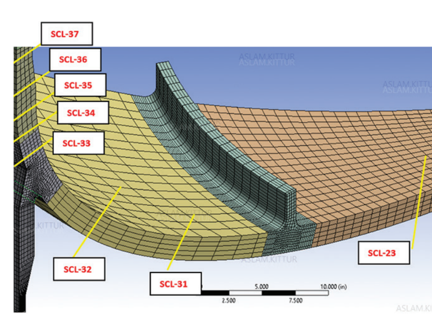

FIG. 7. SCLs in finite element model at bottom head, general.

ties. The stress distribution plots have been provided in **FIGS. 11–13.** The graphical comparison of the simulated values of equivalent stress range (P + Q) at the various SCL locations has been illustrated in **FIG. 14,** whereas **FIG. 15** shows the values of

equivalent stress range (P + Q + F) at the various SCL locations. The maximum accumulated fatigue of 13.5% was calculated at SCL-11 and was found acceptable, as it was within the specified limits. The SCL-11 corresponds to the PSA vessel's skirt section, which was also not a part of the vessel's pressure boundary shell.

The damage factors were calculated based on the stress-fatigue analysis (K_e = 1.0 and K_f = 1.2 were assumed for the calculations), and all SCL locations in the adsorber vessels were found acceptable for continued operation well beyond the previously established fatigue-design life. However, due to embedded flaws, the effect of crack growth has been considered in the following section, which provides the additional criteria to limit the fatigue design life. This shall be based on the crack growth from its size (yr) to its critical size.

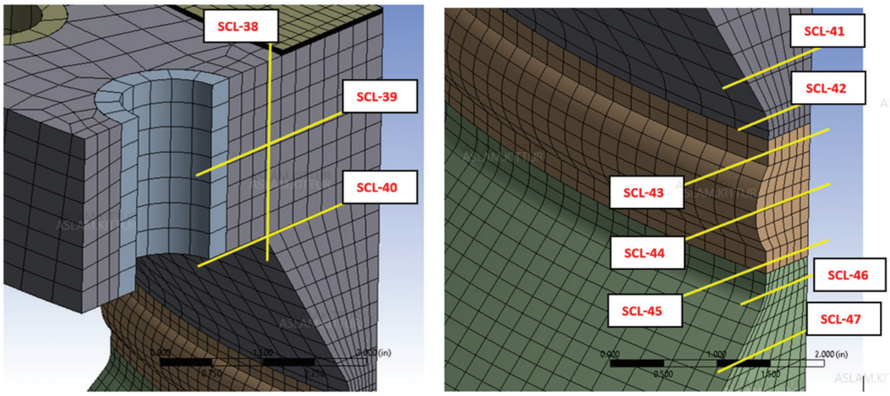

FIG. 8. SCLs in finite element model at top flange.

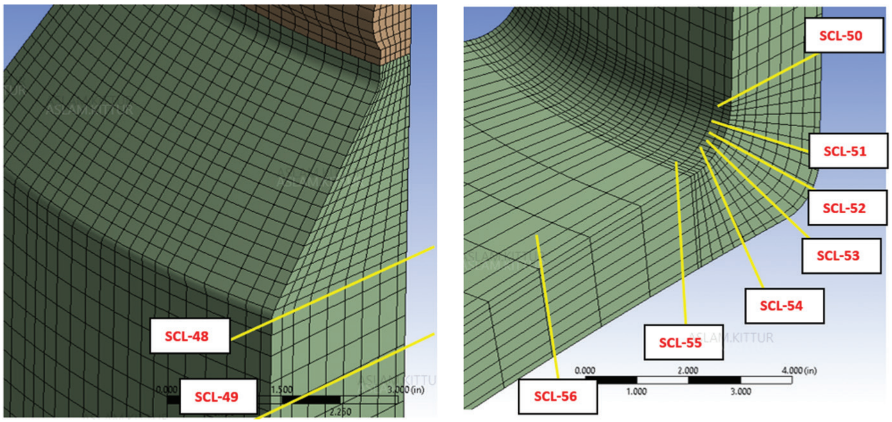

FIG. 9. SCLs in finite element model at top head, local.

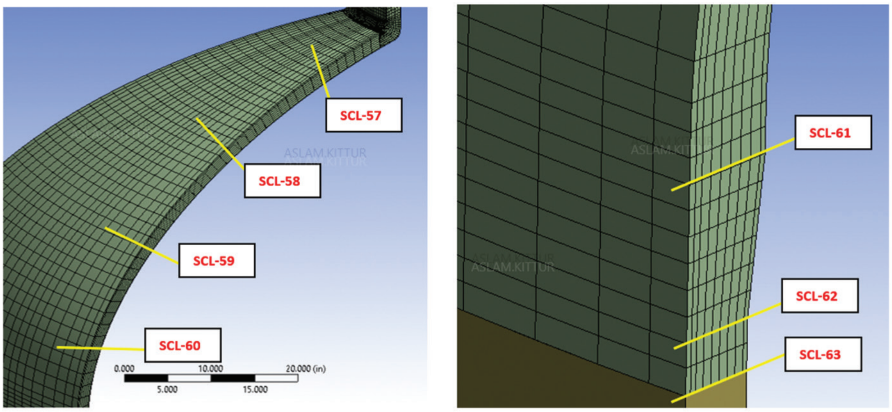

FIG. 10. SCLs in finite element model at top head, general.

Crack criticality assessment. The PSA vessels in pressure cycling service containing the crack-like flaw are subject to loading conditions that may result in crack growth. As a result, the Level 1 and 2 assessment procedures in API-579[1] Part 9 do not apply as conditions are not satisfied. As the embedded flaw is anticipated to be subjected to subcritical crack growth during future operation, it has been evaluated using a Level 3 assessment procedure that provides the best estimate of the structural integrity of a component with a crack-like flaw. A Method A assessment has been utilized, such that the FAD was utilized for the acceptance criteria with user-specified partial safety factors (PSFs) based on a risk assessment.

The flaw characterization rules in API-579[1] will allow existing or postulated crack geometry to be modeled by a geometrically simpler one to make the actual crack geometry more amenable to fracture mechanics analysis. As a result, these linear indications were categorized as embedded cracks [**FIG. 16** (cylinder—embedded crack, longitudinal direction and elliptical shape)] and [**FIG. 17** (cylinder—embedded crack, circumferential direction and elliptical shape)] of API-579[1]. The crack depth, length, angle and distance to other embedded cracks were typically required to be determined using the angle beam ultrasonic testing (UT) examination techniques, such as time of flight diffraction (TOFD) and pulse-echo techniques, as shown in **TABLE 3.**

The Mode I stress intensity factor solution representing a fourth-order polynomial stress distribution (KPECE2) was chosen to represent the cylinder—embedded crack, elliptical shape in the longitudinal direction and circumferential direction. For the FAD, the Level 2 recommended curve was used in the assessment (Level 3).

Stress analysis results from the previous section were used in a crack-like flaw assessment, such that the numerical model did not include the crack explicitly. This approach entails post-processing the stress values in such a way as to tie into existing K_I and σ_{ref} solutions in Annex C and Annex D, respectively. This approach was applied because it is best suited to the case where the PSA vessel's shell could

be approximated by simple shapes, such as a cylinder addressed in Annex C and Annex D of API-579[1]. The stress results were linearized based on the crack location within the section thickness—this linearization method employed K_I estimation, as the stress distribution was uniform. The reference stress σ_{ref} was determined from the membrane and bending stresses derived from the linearization of stress results. The lower-bound estimate of fracture toughness 69.141 ksi√in. used in this assessment was based on the API-579[1] equation F.54 (Eq. 1):

$$K_{IC} = 33.2 + 2.806 \exp\left[0.02(T - T_{ref} + 100)\right] \quad (1)$$

The primary stress, material fracture toughness and flaw size should be modified using the PSF. However, as the conservative estimate of lower-bound fracture toughness has been established, the applicable PSF equal to 1.0 would have normally been permissible. However, for conservatism, a PSF = 1.3 was used in this assessment.

The toughness ratio (K_r) was calculated based on the applied stress intensity due to the primary stress distribution and was plotted as the ordinate of the above FAD assessment point. The load ratio was calculated using the reference stress for primary loads (L^P_r) and plotted as the abscissa of the above FAD assessment point. The FAD plot in **FIG. 18** shows that all existing flaws are well within the acceptable region and were deemed non-critical.

Crack growth assessment. It is anticipated that these cracks may grow by fatigue when the adsorber vessels continue operating, witnessing a PSA pressure cycle that results in cyclic stresses. Each increment of crack extension correlates to a certain increment of stress cycles.

The following basis was adopted for the crack growth assessment methodology:

1. During the crack growth, crack dimensions a and c will grow simultaneously, with the a/c ratio being assumed to remain constant during crack growth of 12.5%.
2. Crack dimension d_1 has been taken from the closest of the two edges, such that $d_1 \leq 0.5$, and is assumed not to change during crack growth.
3. 1st limitation on crack growth (for those cracks that exceed the $a/d_1 \geq 0.8$ criteria before exceeding the cut-off limit for steel): During the crack growth, crack dimension a will increase, resulting in a possible highest $a/d_1 \geq 0.8$. As the crack face nears the thickness boundary, a becomes = d_1. This limiting/highest crack ratio a/d_1 = 0.8 will correspond to the critical size.
4. 2nd limitation on crack growth (for those cracks which exceed the cut-off limit for steel before exceeding $a/d_1 \geq 0.8$ criteria): During the crack growth, crack dimension a will increase, resulting in a FAD position in the unacceptable region. The crack dimension will correspond to the

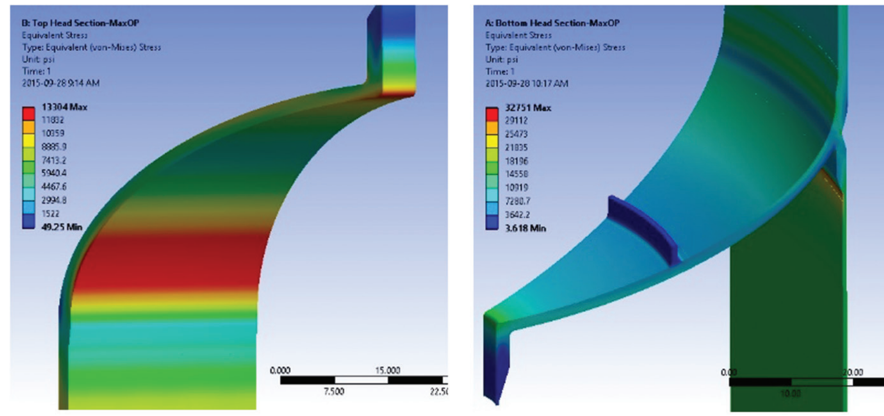

FIG. 11. Stress plots: top and bottom heads.

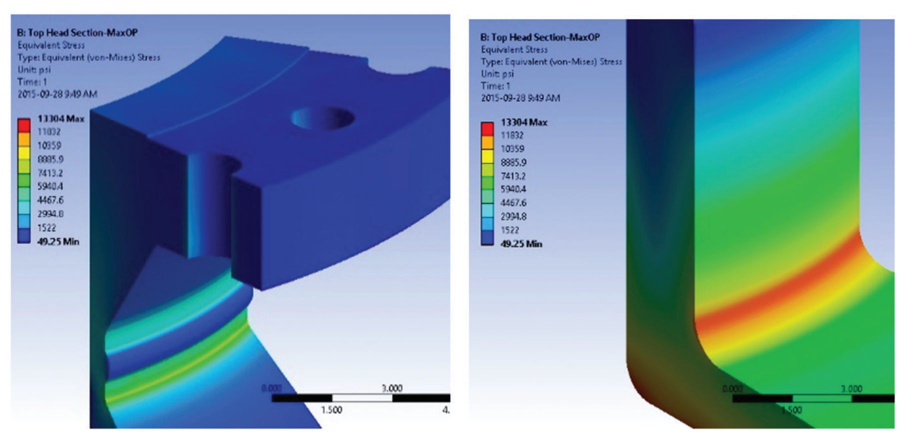

FIG. 12. Stress plots: flange and nozzle neck.

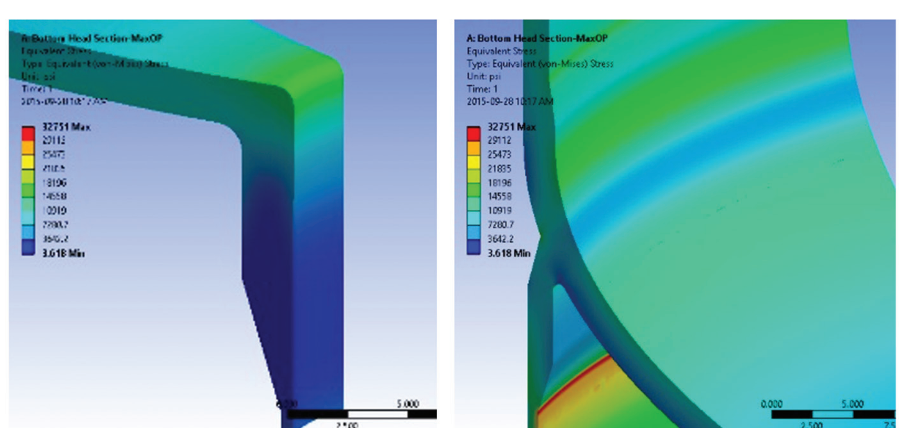

FIG. 13. Stress plots: bottom head.

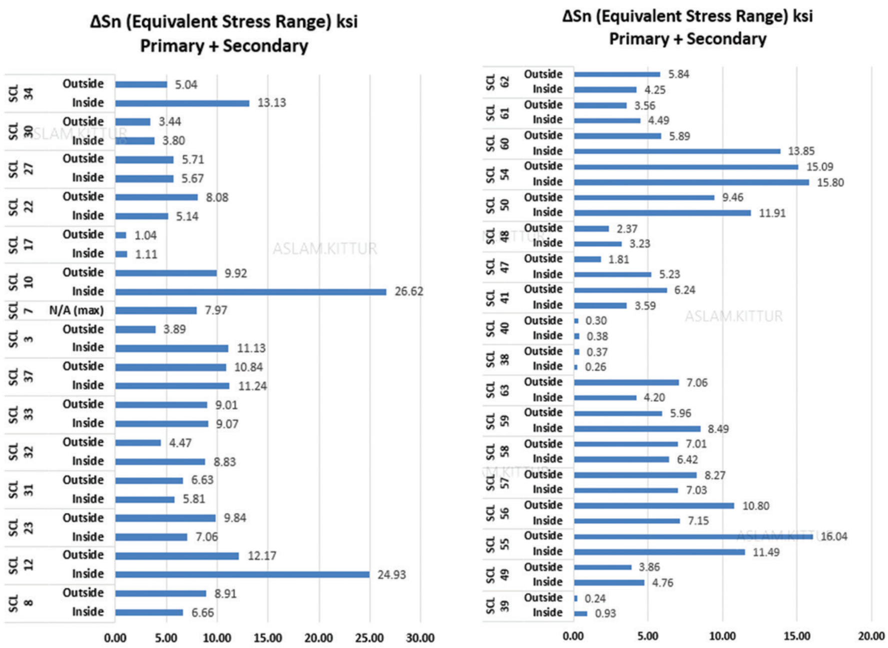

FIG. 14. Equivalent stress range (P + Q).

FIG. 15. Equivalent stress range (P + Q + F).

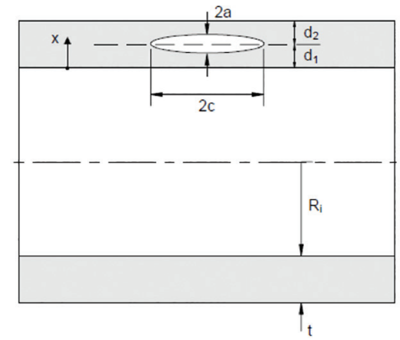

FIG. 16. API-579 (C.19).

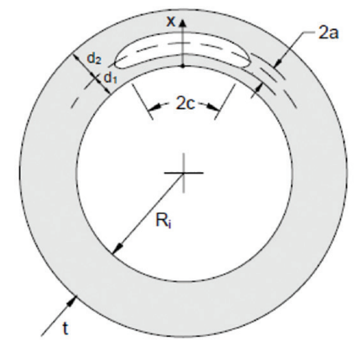

FIG. 17. API-579 (C.20).

crack ratio a/d_1. It is taken as the critical crack size when it touches this limit.

A crack growth model and associated constants were utilized to estimate the remaining life of the adsorber shell locations with the crack-like flaws based on a fracture mechanics approach. The Paris model was chosen for the assessment, as it accounted for environmental effects and was related to cyclic behavior da/dN. The fatigue crack growth equation was used with the Paris Equation (F-1.23) in FFS assessments completed in this analysis, as these equations are valid for materials with yield strengths less than or equal to 600 MPa (87 ksi), shown in Eq. 2:

$$(da/dN) = 8.61(10^{-10})(\Delta K)^{3.0} \quad (2)$$

Where the below threshold stress intensity value (Eq. 3) was used:

$$\Delta K_{th} = 1.8\ ksi\sqrt{in} \quad (3)$$

FIG. 19 provides the graphical illustration of the years required by each recorded crack to reach its respective critical sizes while the concerned PSA continued to operate in the H_2 plant. Following this assessment, the NDE-based inspection of the embedded flaws was completed after a 5-yr gap by entering the PSA vessels in an emptied state during the scheduled catalyst change-out activity. The UT-based TOFD technique was employed to establish the data of any growth of these previously detected flaws during this period. The results confirmed that there was no observable crack growth based on the prior baselined NDE results for characterized geometries of the flaws.

Takeaway. Based on the stress-fatigue analysis in accordance with the requirements of API-579[1], all SCL locations in

TABLE 3. Dimensions of embedded flaws

Flaw location	2a (in.) width	2c (in.) length	d_1 (in.) depth	*t*	0.125 < *a/c* ≤ 1	0.1 ≤ d_1/t ≤ 0.9	*t*	Ratio a/d_1
270D-TC/CS-I1	0.0492125	0.433	1.0236	1.6535	0.114	0.619	1.6535	0.024
270D-TC/CS-I2	0.05905	0.3937	0.2677	1.6535	0.15	0.162	1.6535	0.11
270D-TC/CS-I3	0.0492125	0.4724	0.4724	1.6535	0.104	0.286	1.6535	0.052
270D-TC/CS-I4	0.063975	0.433	0.5118	1.6535	0.148	0.31	1.6535	0.063
270D-BC/LS-I1	0.1033375	0.70866	0.6102	1.6535	0.146	0.369	1.6535	0.085
270D-BC/LS-I2	0.0738125	0.5118	0.748	1.6535	0.144	0.452	1.6535	0.049
260D-BC/PT-01	0.1181	0.3149	0.735	1.6535	0.375	0.445	1.6535	0.08
260D-BC/PT-02	0.0787	0.3937	0.734	1.6535	0.2	0.444	1.6535	0.054
260D-BC/PT-03	0.1181	0.1968	0.797	1.6535	0.6	0.482	1.6535	0.074
260D-BC/PT-04	0.1574	0.1574	0.774	1.6535	1	0.468	1.6535	0.102
260D-BC/PT-05	0.1181	0.1968	0.849	1.6535	0.6	0.513	1.6535	0.07
260D-BC/PT-06	0.1574	0.1181	0.834	1.6535	1.333	0.504	1.6535	0.094
260D-BC/PT-07	0.1968	0.2362	0.803	1.6535	0.833	0.486	1.6535	0.123
260D-BC/PT-08	0.0787	0.3149	0.659	1.6535	0.25	0.399	1.6535	0.06
260D-BC/PT-09	0.0787	0.1574	0.652	1.6535	0.5	0.394	1.6535	0.06
260D-BC/PT-10	0.1181	0.2755	0.782	1.6535	0.429	0.473	1.6535	0.076
260D-BC/PT-11	0.1574	0.1968	0.865	1.6535	0.8	0.523	1.6535	0.091
260D-BC/PT-12	0.1181	0.2362	0.848	1.6535	0.5	0.513	1.6535	0.07

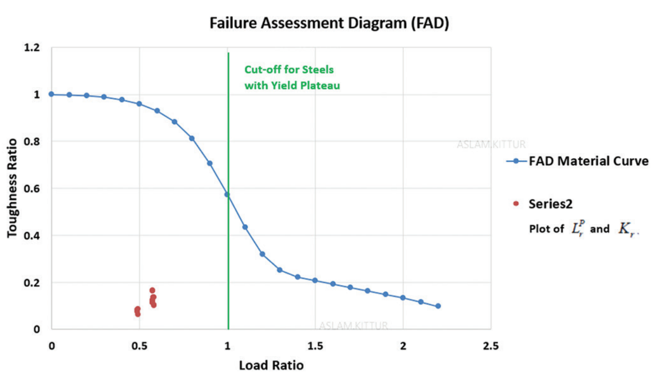

FIG. 18. Fatigue assessment diagram (FAD).

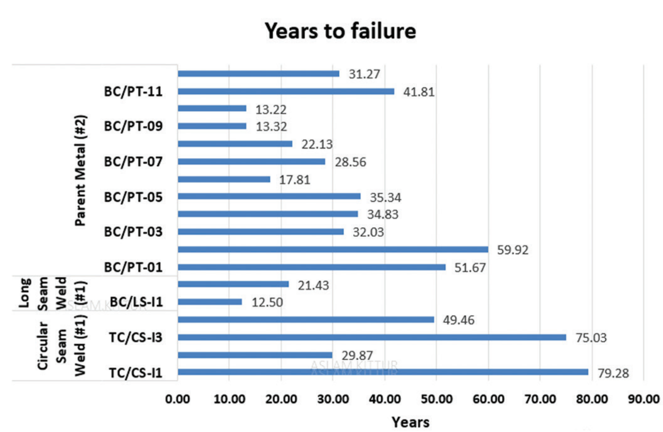

FIG. 19. Fatigue life based on crack growth.

the adsorber vessels were found to be acceptable for continued operation well beyond the previously established fatigue design life. Considering the presence of 18 embedded flaws in two out of the 24 PSA vessels, the least fatigue remaining life of 13 yr was estimated for at least three flaws. On average, all other remaining embedded flaws have a fatigue remaining life of at least an additional 20 yr. To understand that the operational PSA adsorber vessels had been assessed for life extension beyond their originally estimated fatigue design life, it was deemed essential that the embedded flaws be subjected to rigorous NDE-based monitoring to guarantee the continued safe and reliable operation of the PSA vessels. This analysis established that the actual fatigue life of these adsorber vessels exceeded the originally anticipated fatigue life of 20 yr. This had a positive impact on cost projection and lifecycle value analysis of these assets and will also help as the basis for a phased replacement approach of the PSA vessels in the near future. HP

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4 "ANSYS Theory & Application Guide," November 2013.

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